SIMULATION OF FLUID CATALYTIC CRACKER

A thesis submitted in partial fulfillment of the requirements for the degree of

Bachelor of Technology

in

Chemical Engineering

By

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National Institute of Technology

Rourkela

CERTIFICATE

This is to certify that the project report entitle "**SIMULATION OF FLUID CATALYTIC CRACKER**" submitted by **KULDEEP PRADHAN (ROLL NO: 108CH006)** in the partial fulfillment of the requirement for the degree of the B.Tech in Chemical Engineering, National Institute of Technology, Rourkela is an authentic work carried out by him under my super vision. To the best of my knowledge the matter embodied in the report has not been submitted to any other university/institute for any degree.

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ABSTRACT

Fluid catalytic cracking unit (FCCU) performs the most vital role in modern refinery process because it is used for producing more economic refinery products. Crude oil contains hydrocarbons ranging from light gases, LPG and gasoline to residues of high boiling point range. Feed to the FCC unit is the residual product from the distillation column; fluid catalytic cracking (FCC) units convert a portion of the heavy material into lighter products, mainly gasoline, olefins, coke and LPG. Simulation of the fractional distillation was being done to find out the feed composition which is the inlet to the riser reactor. The FCC unit was later simulated to get the final yield of gasoline and other valuable product like LPG and the yield obtained by simulation is acceptable in plant scale. Later different values of flow rate, feed temperature, riser time and temperature of the reactor were varied to get the simulated data and from that graphs were plotted to study the behavior of the reactor and from there optimum conditions for the reactor is concluded. Comparison of single and dual riser is done and optimum condition. Additional unit like fractionator is used in the further simulation which hardly affects the riser reactor behavior. Also the effect of process parameters is studied in the case of dual riser reactor.as per the regenerator CFD simulation is done using ANSYS FLUENT 13.0 to show the temperature profile in the regenerator and the reaction product CO_2. From the CFD analysis the catalyst bed condition and the heat supplying regenerator's profile are described. At various concentration of oxygen enrichment temperature profile is observed and the rise of temperature is noted with high oxygen enrichment. Also rise in flow rate of air to the regenerator affect the combustion on the regenerator and hence the rise in temperature.

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1. Introduction

Catalytic cracking is one of the most important refinery processes in petroleum industries. It has got various units like fluidized catalytic cracking, hydro cracking etc. where heavy oil with higher boiling point is cracked into lighter products i.e. cracked into products of lower boiling point with lower molecular weight like gasoline. According to studies in United States fluidized catalytic cracking process (FCC) provides about 35 to 45% of the blending stocks in refinery of gasoline [1][8]. Before in conventional processes cracking was achieved by thermal cracking process but now it has already been replaced by catalytic process because of its high efficiency and selectivity i.e. gasoline is being produced with higher octane value and less heavy fuel oils and less light gases. The light gases produced in the process contain more olefin hydrocarbons than those by the thermal cracking process [2][8].

The cracking reaction in the catalytic reactor produces coke (carbon), which remains on the surface of the catalyst which decreases the efficiency of the catalyst and its activity decreases. To maintain the activity of the catalyst it is necessary to burn off the deposited carbon on the catalyst. This was done on a regenerator and the active catalyst is further fed back to the reactor. As known, the cracking reaction is endothermic so the energy required for the process comes from the regenerator where catalyst is burned off in presence of air which is an exothermic reaction. Some units like FCC are designed to use the supply of heat from the regenerator for the cracking purpose. These are known as “heat balance “units [3].

The catalytic cracking process can further be classified into 2 major units like the moving bed reactor and the fluidized cracking reactor of which the fluidized cracking reactor has taken over the majority of the production scale now days. The details of FCC units will be discussed later on. But the processes adopted on both the method are almost same.

As discussed heat for the cracking reaction in the riser comes from the burning off the spent catalyst in the regenerator, so the temperature of the regenerator has to be regulated, otherwise overheating of the catalyst or de-selectivity of the process might take place. Regulation of the temperature can be done by maintaining a fixed exit CO_2/CO ratio or the temperature of the regenerator can be fixed accordingly supply of oxygen is controlled.

Fluidized catalytic cracking unit is the primary and the most important conversion unit in the refinery process. Crude oil as obtained from the ground is processed through several separation process like atmospheric distillation column, vacuum distillation column and finally oils of different boiling point ranges are obtained like gasoline (naphtha's), diesel oil, LPG etc. including these products heavy oil (atmospheric gas oil or vacuum gas oil) are produced which has a boiling point 343°C (650 °F) to 565°C (1050 °F). These heavy oils are cracked in the FCC rector to form economically valuable petroleum products like gasoline LPG, lighter olefins. FCC unit is much preferred than the conventional thermal cracking process because it produces petroleum products of higher octane value.

As of 2006, FCC units were in operation at 400 petroleum refineries worldwide and about one-third of the crude oil refined in those refineries is processed in an FCC to produce high-octane gasoline and fuel oils[4]. During 2007, the FCC units in the United States processed a total of 5,300,000 barrels (834,300,000 liters) per day of feedstock [5] and FCC units worldwide processed about twice that amount.

The FCC process employs a catalyst in the form of very fine particles (size of the catalyst is about 70 micrometers (microns)), which behave as a fluid when aerated with vapor. So here the catalyst acts as an agent for both mass transfer operation and heat transfer operation. Catalyst moves from regenerator to reactor and vice versa as fresh or spent catalyst and provide heat to the reactor. Usually two types of FCC units are used in industrial scale which are side by side type and ortho-flow or stacked type reactor. In side by side reactor which will be used in the project for simulation purposes, reactor and regenerator is separated vessel placed side by side. In case of stacked type reactor rector and regenerator are mounted together, the later mounted above before.

The basic process of FCC has got two major components i.e. reactor and regenerator. All the major processes happen here which are divided into following categories.

1.1. Preheat system

The residue and the Atmospheric gas oil from the distillation column are two major components of feed for the FCC reactor. These feed are to be preheated before entering into the reactor. This is done by the feed preheat system which heats both the fresh and recycled feed .pre-heating is done through several heat exchangers and the temperature maintained is about 500-700 °F

1.2. Reactor

Until about 1965, units were designed with a dense phase fluidized bed in the reactor vessel. The units were modeled and also operated so that all the reaction occurs in the reactor section. Now it has been developed that majority of the reaction occurs in the riser as the catalyst activity and temperature were at their highest there. No significant attempts were made for controlling the riser operation. But after the usage of the reactive zeolite catalyst the amount of cracking occurring in the riser has been enhanced. Now the reactor is used for the separation purpose of both the catalyst and outlet products. Reaction in the riser is optimized by increasing the regenerated catalyst velocity to a desired value in the riser reactor and injecting the feed into the riser through spray nozzles.

The fresh feed and the recycled streams are preheated by heat exchangers or a furnace and then enter to the riser where they were mixed with the hot regenerated catalyst. The heat from the catalyst vaporizes the feed and required temperature for the reactor has been attained. The mixture of catalyst and hydrocarbon vapor travels up the riser into the separator. The cracking reaction starts when the feed is in contact with the hot catalyst in the riser and continues until oil vapors are separated from the catalyst in the reactor separator. The hydrocarbons are then sent to the fractionator for the separation of liquid and the gaseous products. In the reactor the catalyst to oil ratio has to be maintained

Properly because it changes the selectivity of the product .the catalyst sensible heat is not only used for the cracking reaction but also for the vaporization of the feed. The ideal riser diameter would be about 2 meters and length is about 30 to 35 meters. During simulation the effect of the riser is presumed as plug flow reactor where there is minimal back mixing, but practically there is both downward and upward slip due to drag force of vapor. [6][9]

1.3. Regenerator

The catalyst comes to the reactor through the catalyst stripper. Regenerator maintains the activity of the catalyst and also supplies heat to the reactor. Depending upon the feed stock quality there is deposition of coke above the catalyst surface. To reactivate the catalyst air is supplied to the regenerator by using large air blower. High speed of air is maintained in the regenerator to keep the catalyst bed in the fluidized state. Then through the distributor at the bottom air is sent to the regenerator .coke is burned off during the process .the heat is produced due to the combustion of the coke and this heat is utilized in the catalytic cracking process. Heat is carried by the catalyst as

sensible heat to the reactor. Flue gas coming out of the regenerator is passed through the cyclone separator and the residual catalyst is recovered. The specification of the catalyst will be discussed in detail at literature review. The regenerator is designed and modeled for burning the coke into carbon monoxide or carbon dioxide. Before conversion of carbon to carbon monoxide was done as half of the air supply required for the process so that the capital cost will be minimum but now a days air is supplied in such scale that carbon is converted into carbon dioxide .in this case the capital cost will be higher but the regenerated catalyst will have a minimum coke content on it. This gives a more efficient and selective catalyst in the riser.

1.4. Flue gas system

It is the heat recovery system of the FCC unit. The flue gases obtained due to burning off the carbon contains a lot of energy which is recovered in this unit. The flue gases like carbon monoxide are burned off in a carbon monoxide furnace (waste heat boiler) to carbon dioxide and the available energy is recovered. The hot gases can be used to generate steam or to power expansion turbines to compress the regeneration air and generate power.

1.5. Catalyst handling

Catalyst particles lower than 20 microns escape during the burning process as air with high velocity is supplied. The catalyst escaping from the regenerator was stopped and controlled by electrostatic precipitator. It screens the escaping catalyst and sends it back to the regenerator. Like this little bit amount of catalyst escape from the reactor which is collected at the bottom of the fractionator tower

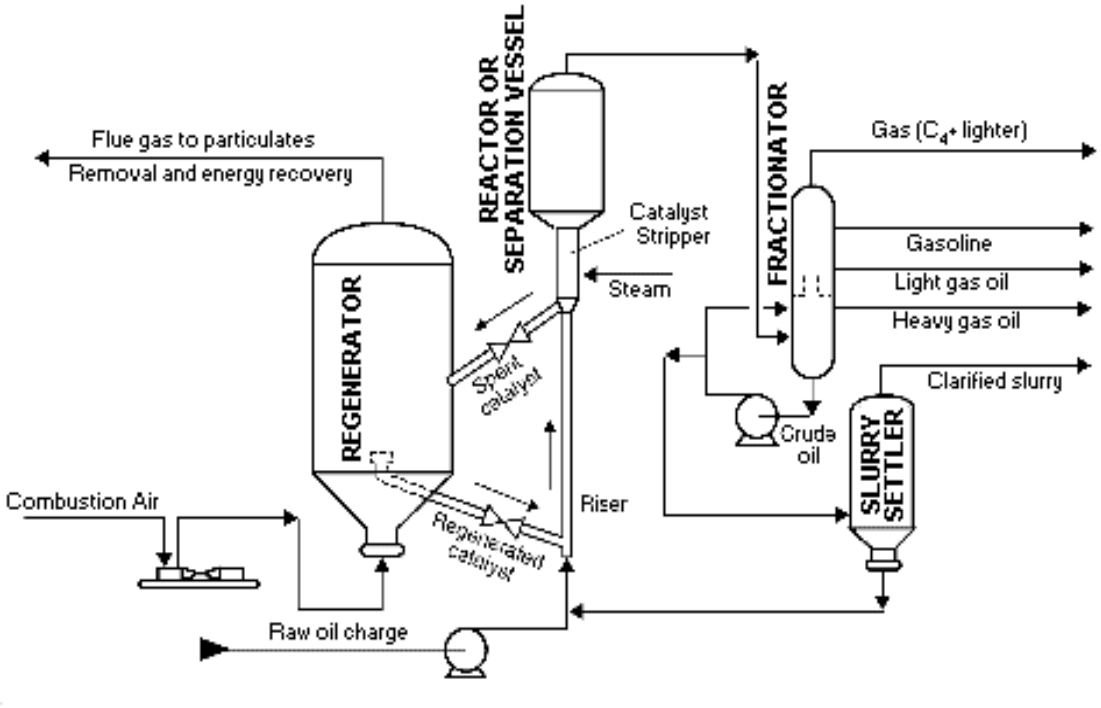

Figure 1: Schematic of the Fluid Catalytic Cracking Unit

So after studying various cases and process parameters simulation of the FCC reactor is done which is the objective of the project. The process parameters are varied at different condition and the efficiency of the reactor was calculated. Simulation is done using Aspen hysys .In the simulation the feed condition was obtained by simulating the atmospheric distillation column.

2. Literature Review

2.1. Reactions in FCC

A series of complex reactions takes place in the riser of the fluidized catalytic cracking unit when feed oil(AGO) is sent to the it with catalyst having the temperature range of about 800^0 F-1200^0F.both catalytic and thermal cracking occur in the riser reactor ; the later one is caused by the non-ideal mixing of the components in the riser. Whether catalytic cracking or thermal cracking, in all types of cracking process C-C bond between the molecules break to give smaller molecules. Before the development of the catalytic process thermal cracking was used , which is a function of time and temperature, the mechanism of which can be explained by free radical formation. Feed oil is exposed to high temperature range of about 1200^0F. Now a day catalytic cracking is accepted everywhere.

The classification of catalytic reactions can be done in two broader sections i.e. primary cracking of gas molecules and secondary rearrangement, re-cracking of the cracked products. The mechanism of catalytic cracking reactions is governed by formation of carbocation, which is of two types' carbonium ion and carbenium ion. A carbonium ion (CH_3^+) is formed when a proton is attacked to the paraffin molecule where as a carbenium ion is formed by adding a positive charge to olefins. Cracking of C -C bond, isomerization reaction, hydrogen transfer reactions are occurred by carbenium ion mechanism .All the cracking reaction occur by beta fission method, because the removal of carbon carbon bond from the beta position requires less energy as compared to the alpha fission.[18]

(1) Catalytic reactions:

(a) Paraffins cracking to olefin and smaller paraffins

$$C_{10}H_{22} \longrightarrow C_3H_6 + C_7H_{16}$$

(b) Olefins cracking to smaller olefins

$$C_8H_{16} \longrightarrow C_5H_{10} + C_3H_6$$

(c) Aromatics side chain reactions

$$Ar - C_{10}H_{21} \longrightarrow Ar - C_5H_9 + C_5H_{12}$$

(d) Napthenes cracking to olefins and smaller rings

$$Cyclo - C_{10}H_{20} \longrightarrow C_6H_{12} + C_4H_8$$

(2) Isomerization reactions

(a) Normal olefins to iso-olefins

$$\text{Iso} - \text{C}_4\text{H}_8 \longrightarrow Trans - C_4H_8$$

(b) Normal paraffins to iso-paraffins

$$\text{n} - \text{C}_4\text{H}_{10} \longrightarrow Iso - C_4H_{10}$$

(3) Hydrogen transfer reactions

$$\text{Napthene} + \text{Olefins} \longrightarrow \text{Aromatics} + \text{Paraffins}$$

(a) Cyclo aromatization

$$\text{C}_6\text{H}_{12} + 3\text{C}_5\text{H}_{10} \longrightarrow C_6H_6 + 3C_5H_{12}$$

(b) Olefins to paraffins and aromatics

$$4\text{C}_6\text{H}_{12} \longrightarrow 3C_6H_{14} + C_6H_6$$

(4) Trans-alkylation / Alkyl group transfer

$$\text{C}_6\text{H}_4(\text{CH3})_2 + \text{C}_6\text{H}_6 \longrightarrow 2C_6H_5CH_3$$

(5) Cyclisation of olefins to Napthenes

$$\text{C}_7\text{H}_{14} \longrightarrow CH\ cycle\ C_6H_{11}$$

(6) Dehydrogenation reactions

$$\text{n} - \text{C}_8\text{H}_{18} \longrightarrow C_8H_{16} + H_2$$

(7) De-alkylation

$$\text{Iso} - \text{C}_3\text{H}_7 - \text{C}_6\text{H}_5 \longrightarrow C_6H_6 + C_3H_6$$

(8) Condensation

$$\text{Ar} - \text{CH} = \text{CH}_2 + \text{R}_1\text{CH} = \text{CHR}_2 \longrightarrow Ar - Ar + 2H$$

Hydrogen transfer reaction (hydride transfer) is carried out as olefins being the main reactant. Two olefins or olefins and naphthenes are reacted in the active site of the catalyst forming paraffin's and cyclo-olefins. Cyclo-olefin is then further converted to paraffin and cyclodi-olefins.

2.2. Pseudo-components

The average °API is estimated from the °API curve of the crude and the TBP of the crude/product. This estimation is useful in evaluating the mass balances from volume balances. Oil refinery processes are usually reported in terms of barrels, volumetric flow rate. The volumetric flow rate can be converted to mass flow rate through the use of average °API of the stream obtained through this process.

The concept of pseudo-components is used to estimate the average °API of the crude/product stream by characterizing the TBP curve of the crude/product. Crude oil constitutes about a million compounds or even more, so the pseudo-components concept is utilized to aid refinery process as the refinery process stream could not be represented using 50-100 components. The pseudo-component representation of the crude stream and its subsequent processes is done by characterizing the crude oil into 30-40 pseudo-components whose average properties can be used to represent the TBP, °API and sulfur content of the streams. A pseudo-component is defined as a component that can represent the average mid volume boiling point and its average properties such as °API and percentage sulphur content.

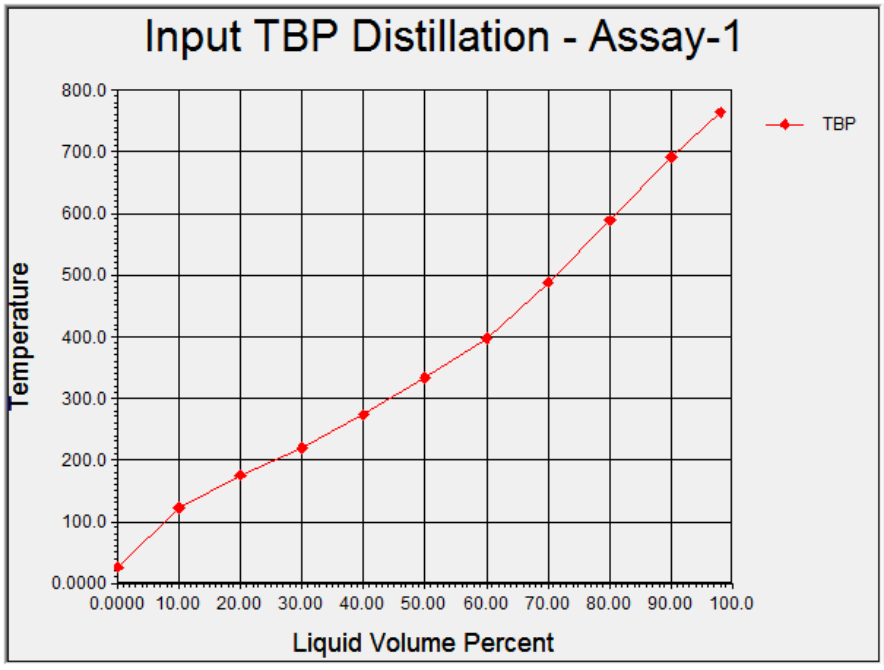

Figure 2: Boiling Temperature vs. Liq. Vol. % for pseudo components

In a typical TBP, a pseudo-component is so created such that within a given range of volume percentage, the pseudo-component covers equal areas under and above the curves (Figure 2). This is possible if the area for the volume cuts corresponds to a straight line, with the fact that for this straight line cutting exactly at the mid-point, the areas above the straight line and below the straight line are truly equal. In order to represent this non-linear curve, a large number of straight lines are necessary making the calculation procedure tedious. Hence, a crude/product stream typically is represented by 20-30 pseudo-components. Corresponding to the pseudo-component, the temperature to represent a section of the crude volume on the TBP is termed as mid boiling point (MBP) and the volume as mid volume (MV). Each pseudo-component has cut points, a temperature range for the pseudo-components. The TBP can then be converted into a tabulated form comprising of hypo-component number, section temperature range, section volume range, MBP and MV.

2.3. Catalytic activity

Commercial catalyst for the cracking reactions are of 3 types; 1) acid treated natural aluminosilicates, 2) amorphous synthetic silica alumina combinations and 3) crystalline synthetic silica alumina catalysts called zeolites or molecular sieve.[7]now a days for getting the optimum production mixture of 2 & 3 or third type of catalysts are used.

The typical FCC catalyst consists of a mixture of an inert matrix (kaolin), an active matrix (alumina), a binder (silica or silica–alumina) and a Y zeolite. During the FCC process, a significant portion of the feedstock is converted into coke. [10]. For the selectivity of the product zeolite is the essential part which ranges about 15 to 25 % of the catalyst and its structure is like tetrahedron with four oxygen at the corner having an aluminum or silicon at the center. In general, the zeolite does not accept molecules larger than 8 to 10 nm to enter the lattice [11].

Y-zeolite is the active and the most important component in FCC catalysts. It provides the major part of the surface area and the active sites [12.]. Thus, it is the key component, which controls catalyst activity and selectivity [13]. The catalytic activity of Y-zeolite is mainly controlled by its unit cell size (UCS) and to less extent by its crystal size. Recently, Al-Khattaf and de Lasa have studied the effect of Y-zeolite crystal size on the activity and selectivity of FCC catalysts[14.] [15]. The conversion of coke and other catalytic activity depends on the acidic strength of the zeolite. So it is known that increase in the yield of coke occurs when there is high acidic strength (high UCS) value .high UCS also favors the hydrogen transfer reaction. As it is discussed the coke

yield increases due to high UCS and it covers the active acidic part of the catalyst which decays the activity. More over the concept of octane number plays a vital part in selectivity of the reactor. That is why hydrogen transfer reaction an important one in the catalytic cracking reactor as it converts some of the light olefins into paraffin's and aromatic compounds which have higher octane number value. [16] .As strong acidic strength is needed for the cracking and a little bit weak acidic condition favors the reaction process of hydrogen transfer so blending or modification of the catalyst is needed accordingly so that the product obtained would be economically valuable. Also the UCS of the catalyst decides the yield percentage of gasoline in the reactor.

Three types of USY zeolite (TSZ-330HSA,TSZ-330HUA, TSZ-360HUA) supplied by Tosoh Corporation and an NH4-Y zeolite (LZY 62) supplied by Union Carbide Corporation weremixed with kaolin clay and stabilized silica sol (conc .30%, 50% ,20% in order) and then some sample calcinated and some steamed at various cond. given below to form catalyst.

Table 1: Cracking composition by using zeolite of Tosoh Corporation and Union Carbide Corporation[15]

Catalyst	**Treatment(°C)**	**UCS**	**ACID**
CA600	Calcination (600)	24.264	0.044
CE500	Calcination (600)	24.577	0.239
CB600	Calcination (600)	24.363	0.0819
SB710	Steaming (710)	24.281	0.0474
SB760	Streaming (760)	24.244	0.0293
CC600	Calcination (600)	24.401	0.121
SC810	Steaming (810)	24.206	0.019
SC600	Steaming (600)	24.317	0.0868

Now selectivity is studied in accordance to the above experiment. As the coke yield will be maximum if there are dense active acidic sites and so it will be more if maximum cracking takes place .Also the coke conversion is directly proportional to the UCS. Now in the case of gasoline production there is always the danger of over-cracking where gasoline productivity decreases .as due to over-cracking on very high UCS value gasoline yield decreases. High UCS values are appreciated in petrochemical industries where light olefins are the major product.

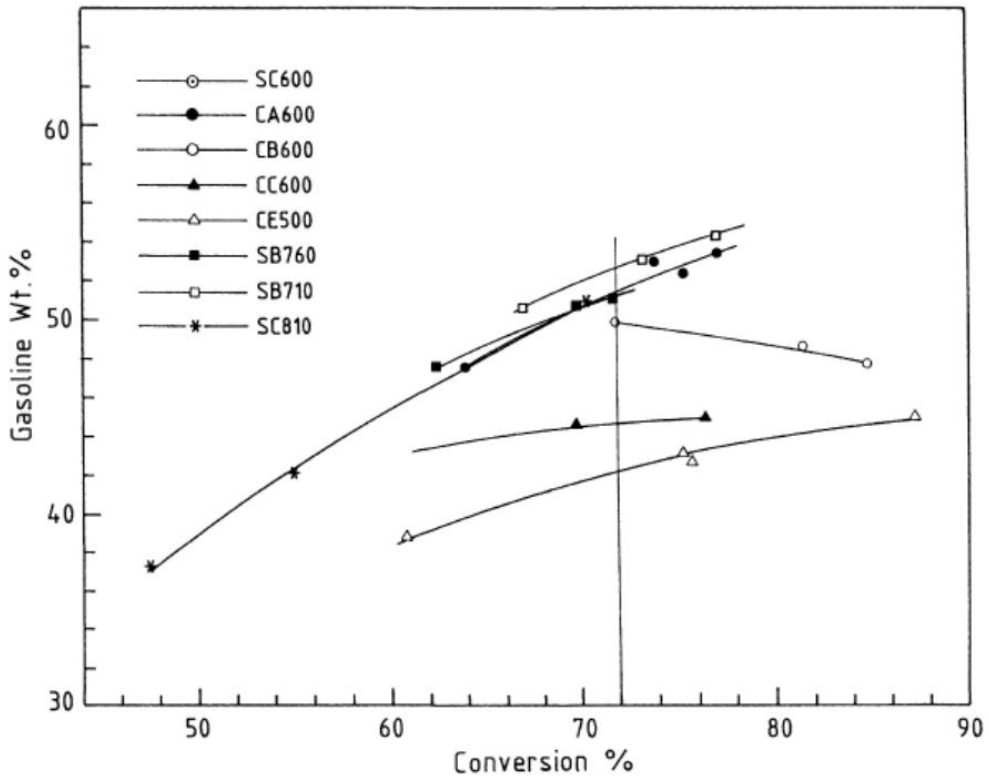

Figure 3: Conversion % vs. Gasoline wt. % for different catalyst

3. Description of the Simulation

ASPEN HYSYS simulator provides an opportunity to check the feasibility of a process, to study and investigate the effect of various operating parameters on various reactions. It is a strong tool for simulation studies and helps in analyzing the outcome of a process. HYSYS offers a high degree of flexibility because there are multiple ways to accomplish specific tasks. This flexibility combined with a consistent and logical approach to how these capabilities are delivered makes HYSYS an extremely versatile process simulation tool. The usability of HYSYS is attributed to the following four key aspects of its design:

• Event Driven operation
• Modular Operations
• Multi-flowsheet Architecture
• Object Oriented Design

3.1. Problem Description

The effects of various operating and design conditions were to be tested and its effects on naphtha yield, coke yield and total conversion were to be noted.

Preheat crude passed through flash evaporator forming pre flash vapor and pre flash liquid which is later passed through heater. Both the components passed through mixer and then to the fractionating column. Atmospheric gas oil goes to the FCC Unit for cracking. Further description is given below.

3.2. Aspen Hysys Simulation

The FCC unit works through various cracking reaction in parallel in the riser reactor section of this unit. Different types of model FCC reactors are available in ASPEN HYSYS such as:

1. One riser
2. Two riser
3. Risers with fractionator

In order to process and separate desired products several unit operations such as fractionation column and product blender could also be used.

In order to operate the FCC unit the feed input to the unit is required which is why entire process of distillation was done. Various components were used to perform the process.

They are:

1. A Separator (flash process) separating into pre-flash liquid and pre-flash vapor of temperature 450°F.
2. A Heater for heating the pre flash liquid to 650°F.
3. Mixer was used to mix these above 2 components to provide a heated feed for the Atmospheric Distillation Tower.
4. Atmospheric Distillation Tower with 29 trays was used having 3 side strippers having 3 stages each (total 40 stages including reboiler and condenser).
5. 3 coolers for the Atmospheric Distillation Tower.

3.3. Simulation

The main purpose of the project includes the effect of variation of process conditions on the production of naphtha yield in the FCC. For the present study, a refinery process was simulated in order to assist in the simulation. The details are discussed below:

3.3.1. Process Flow Diagram

To represent the refinery process + FCC unit in Aspen HYSYS, the first step is to make a process flow diagram (PFD). In Simulation Basic Manager, a fluid package was selected along with the components which are to be in the input stream. In the process, Peng-Robinson was selected as the fluid package as it is able to handle hypothetical components (pseudo-components).

The non-oil components used for the process were H_20, C3, i-C4, n-C4, i-C5 and n-C5. The pseudo-components were created by supplying the data to define the assay. The fluid package contains 44 components (NC: 44): 6 pure components (H_2O plus five Light Ends components) and 38 petroleum pseudo-components). In order to go to the PFD screen of the process the option "Enter to simulation Environment" was clicked on. An object palette appeared at right hand side of the screen displaying various operations and units.

The PFD of the process is given below:

Where,

PreFlash is a separator.

Furnace is a heater.

Mixer is a mixer.

Atmos. Tower is a distillation column operated at 1 atm.

Reactor Section is the FCC Unit in which AGO (Atmospheric Gas Oil) is used as the feed.

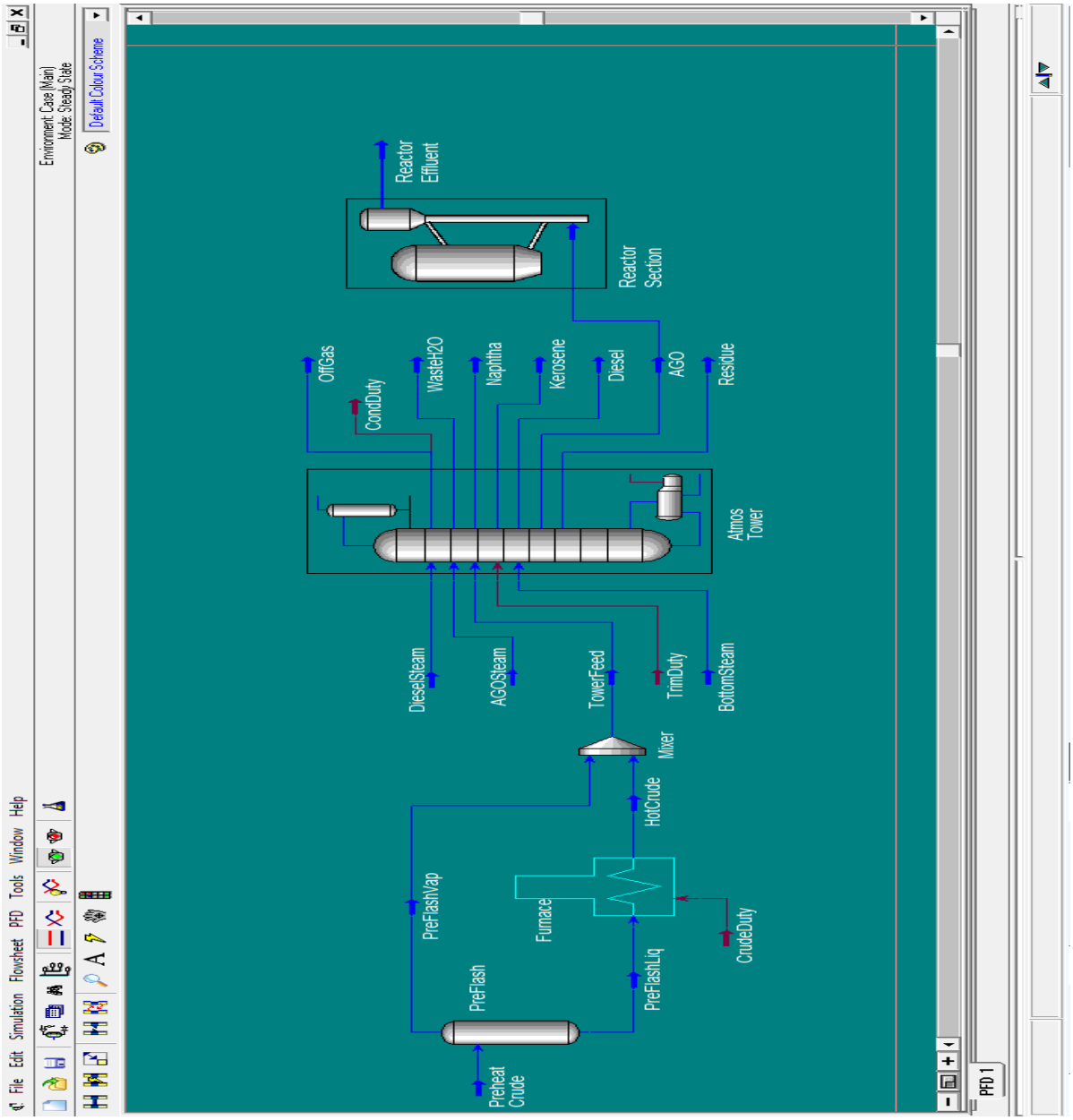

Figure 4: PFD of the simulation carried out in ASPEN HYSYS

3.3.2. The Process

A Crude Oil enters the PreFlash unit, a separator used to split the feed stream into its liquid and vapor phases at 450^0 F and 75 psia having a molecular weight of 300 and °API of 48.75. The crude stream separates into the PreFlashVap and PreFlashLiq consisting of purely vapor and liquid respectively. The PreFlash Liq enters the crude furnace flashing part of the liquid to vapor which comes out as stream, Hot Crude having a temperature of 650^0 F. The Pre-FlashVap.And Hot Crude streams are then inlet into the Mixer resulting into the formation of the Tower Feed. The Atmos. Tower is a column having Side Stripper systems to draw out Kerosene, Diesel and Atmospheric Gas Oil. Naphtha is drawn from the condenser and Residue from the reboiler. The Atmospheric Gas Oil (AGO) is then used as the feed to the Reactor Section, the FCC unit. The FCC Unit was configured to have one or two risers with the geometry as per the data collected by Ali et al [17]. It was assumed that no heat loss occurs in the FCC unit. Catalyst was decided upon and operating conditions were set.

Results were noted for the variation of Naphtha Yield, Coke (wt. %) and Total conversion with change in the following operating conditions:

i) C/O ratio
ii) Feed Flow Rate
iii) Feed Temperature
iv) Reactor Temperature
v) Reactor Pressure

Total conversion is attributed to the conversion of the feedstock to the FCC into H_2S, Fuel Gas, Propane, Propylene, n-Butane, i-Butane, Naphtha, Butenes and Coke while the conversion of feedstock to Light Cycle Oil and Bottoms is not considered in the calculation of total conversion.

3.3.3. The components or the blocks or the equipment's

Description of various components used in the PFD and the conditions at which they are operated are described here:

a) Separator (PreFlash)

No heat loss was assumed for the separator of volume 70.63 ft^3. Preheat Crude entered at 450^0 F and 75 psia with a 100,000 barrels/day flow rate containing mostly liquid. It had a molecular weight

of 300 and API Gravity of 48.75. The Preheat Crude was separated into PreFlashLiq (450^0 F, 75 psia) and PreFlashVap (450^0 F, 75 psia).

b) Heater (Furnace)

No heat loss was assumed for the Heater. PreFlashLiq entered the furnace at 450^0 F and 75 psia. Its main purpose was to partially vaporize the feed and increase its temperature to the feed conditions needed for the distillation column. The outlet stream Hot crude had conditions 650^0 F, 65 psia.

c) Mixer (Mixer)

The main purpose of the Mixer was to mix two streams, HotCrude (650^0F, 65 psia) and PreFlashVap ($450\ ^0$F, 75 psia) to give on stream, TowerFeed (641.5^0 F, 65 psia) which is the feed stock to the distillation column.

d) Distillation Column (Atmos Tower)

The feed to the column enters at 641.5^0 F, 65 psia. The column separates the feed into six fractions namely: Off Gas, Naphtha, Kerosene, Diesel, Atmospheric Gas Oil and Residue. The main column consists of 29 trays.

e) Fluidized Catalytic Cracking Unit (Reactor Section)

The AGO IS taken as the feed for the fluidized catalytic cracking unit. Initial conditions are given in the appendix attached. Results are shown in the Results and Discussion section.

The simulation for the FCC unit needs simulated feedstock. For the feedstock for the FCCU, Crude Petroleum, data was obtained from ASPEN HYSYS. The feed of molecular weight 300 and API Gravity 48.75 was used at a temperature of 450 °F and pressure of 75 psia.
Given below are the properties used for the crude petroleum feedstock:

Table 2: Crude Petroleum Simulation Feedstock Properties

Preheat Crude (Feedstock)	
Temperature [°F]	450
Pressure [psia]	75
Liquid Volume Flow [gallons/day]	3150000

Table 3: Bulk Crude Properties

Bulk Crude Properties	
MW	300.00
API Gravity	48.75

Table 4: Light Ends Liquid Volume Percent of Crude Petroleum Feedstock

Light Ends Liquid Volume Percent	
i-Butane	0.19
n-Butane	0.11
i-Pentane	0.37
n-Pentane	0.46

Table 5: API Gravity Assay of Crude Petroleum Feedstock

API Gravity Assay	
Liq Vol% Distilled	**API Gravity**
13.0	63.28
33.0	54.86
57.0	45.91
74.0	38.21
91.0	26.01

Table 6: Viscosity Assay of Crude Petroleum Feedstock

Viscosity Assay		
Liquid Volume Percent Distilled	**Viscosity (cP) 100°F**	**Viscosity (cP) 210°F**
10.0	0.20	0.10
30.0	0.75	0.30
50.0	4.20	0.80
70.0	39.00	7.50
90.0	600.00	122.30

Table 7: TBP Distillation Assay of Crude Petroleum Feedstock

TBP Distillation Assay		
Liquid Volume Percent Distilled	**Temperature (°F)**	**Molecular Weight**
0.0	80.0	68.0
10.0	255.0	119.0
20.0	349.0	150.0
30.0	430.0	182.0
40.0	527.0	225.0
50.0	635.0	282.0
60.0	751.0	350.0
70.0	915.0	456.0
80.0	1095.0	585.0
90.0	1277.0	713.0
98.0	1410.0	838.0

The feed was simulated through the process explained above and the product properties for the Atmospheric Distillation Tower were obtained. The Distillation Tower had six outlets out of which the top gaseous product stream had no mass flow. Hence only properties for the five outlet streams which consisted of Naphtha, Kerosene, Diesel, Atmospheric Gas Oil (AGO) and Residue were obtained. The AGO stream was then used in a 1-riser FCC unit to obtain the Naphtha Weight percentage and total conversion by varying different parameters such as reactor temperature and mass flow rate. The conditions under which the FCC unit was operated are given in Appendix 1.

Table 8: Atmospheric Distillation Tower Product Properties

Atmospheric Distillation Tower Product Properties					
Product Name	**Liquid Volume Flow [gallons/day]**	**Molecular Weight**	**Mass Density [API]**	**Temperature [°F]**	**Pressure [psia]**
Naphtha	630000	138.4	86.12	163.9	19.7
Kerosene	409500	210.1	118.8	449.2	29.84
Diesel	535437	289.1	109.6	478.4	30.99
AGO	158035.5	390.1	114.6	567.2	31.7
Residue	1301643	614.6	83.21	657.1	32.7

3.3.4. Regenerator Simulation

As per now only the riser reactor simulation process has been discussed. As no important reaction occurs in the regenerator (only the combustion reaction for the heat supply to the riser.), so HYSYS does not give any profile about it. That's why ANSYS FLUENT is used to conduct the temperature profile as well as the combustion reaction profile. As the process reaction is a multiphase reaction so eulerian-eulerian principle is used. Some model equations are used in the simulation, those are

Continuity equation:

$$\frac{\partial \rho}{\partial t} + \nabla . (\rho \vec{v}) = S_m$$

$\rho =$ Density of the fluid.

$\vec{v} =$ Velocity of the flow.

$S_m =$ The mass added to the continuous phase from the dispersed second phase.

This is general form of mass conservation equation which is valid for both incompressible and compressible flows.

Momentum conservation equation:

Conservation of momentum in an inertial (non-accelerating) reference frame is described by

$$\frac{\partial}{\partial t}(\rho\vec{v}) + \nabla.(\rho\vec{v}\vec{v}) = -\nabla p + \nabla.(\bar{\bar{\tau}}) + \rho\vec{g} + \vec{F}$$

$p =$ The static pressure,

$\bar{\bar{\tau}}$ = the stress tensor (described below), and

$\rho\vec{g}$ And $\vec{F}$ are the gravitational body force and external body forces (e.g., that arise from interaction with the dispersed phase), respectively.

$\vec{F} =$ Contains other model-dependent source terms such as porous-media and user-defined sources.

The stress tensor $\bar{\bar{\tau}}$ is given by

$$\bar{\bar{\tau}} = \mu[(\nabla\vec{v} + \nabla\vec{v}^T) - \frac{2}{3}\nabla.\vec{v}I]$$

Species transport equation:

To Solve conservation equations for chemical species, ANSYS FLUENT predicts the local mass fraction of each species, Y_i, through the solution of a convection diffusion equation for the ith species. This conservation equation takes the following general form:

$$\frac{\partial}{\partial t}(\rho Y_i) + \nabla.(\rho\vec{v}Y_i) = -\nabla.\vec{J_i} + R_i + S_i$$

R_i =the net rate of production of species I by chemical reaction

S_i =the rate of creation by addition from the dispersed phase plus any user defined sources.

$\vec{J}_i$ =diffusion flux of species i, which arises due to gradients of concentration and temperature.

Regenerator is made on workbench having the dimension of diameter 25 ft. and total height 78 ft. Other dimension are taken accordingly .while solving, pressure based solver is used for incompressible fluid. Gravity is defined as -9.81m/s2.Operating pressure is outlet atmospheric and inlet condition is velocity .pressure velocity equation coupling is done by using phase coupled scheme.

Momentum equation is solved by first order upwind difference scheme. For multiphase and species transport, solving momentum equation is done by upwind difference scheme. For solving first order transient scheme is taken (unsteady state). Relaxation factors used for simulation are defined below:

Parameter	Relaxation factor
Pressure	0.7
Momentum	0.3
Phase1 (O_2)	0.5
Phase2	0.5

Then boundary condition and initial conditions are defined. In boundary condition , At the wall no slip condition is applied whereas inlet is the pressure inlet condition and outlet is the pressure outlet. Initial condition is then applied where the velocity inlet temperature is defined as 475K .Also the carbon and catalyst initial temperature is taken 522^0C(795K) according to Ali et al. Initial height of the catalyst bed .volume fraction of alumina is taken as 0.5 and carbon as 0.05 and the rest is void fraction.

4. Results and Discussion:

On simulation of the FCC unit under the above stated condition the following outputs have been obtained giving data on the yield in terms of weight %. The model parameters used for the simulation process comes from Ali et al case study. [17]

Height: 33 meter.
Diameter: 8 meter.
Flow rate: 20kg/sec.
Catalyst to oil ratio: 7.2
Feed temperature: 495.8K
Catalyst used in the process: AF3

Table 9: Outlet Composition Results from FCC simulation

COMPONENTS	**WEIGHT (%)**
H_2S	1.6218
FUEL GAS	7.8644
PROPANE	4.5558
PROPYLENE	7.2727
N-BUTANE	1.9853
I-BUTANE	4.9022
NAPHTHA	39.485
BUTENES	8.9245
LCO	10.1014
BOTTOMS	6.9333
COKE YIELD	6.3534
CONVERSION	82.9653
TOTAL	100

4.1. Effects Of C/O Ratio

Changing catalyst to oil ratio (C/O ratio) in the FCC reactor does remarkable effect on gasoline and coke yields. Though it is not a direct parameter but it regulates the temperature in the riser. The naphtha yield increases with the increasing C/O ratio however, the rate of increase in the naphtha yield decreases at higher values of C/O ratio. This can be known by the fact that at a very high catalyst concentration cracking of pseudo components in the naphtha range (known as secondary cracking reactions) increases which causes a decrease in the rate of increase of naphtha yield with C/O ratio. On the other hand, the increasing C/O ratio leads to increase in catalyst concentration, and hence increase in rate of both primary and secondary cracking. This increases overall number of moles cracked on the catalyst surface and hence increases amount of coke deposited on the catalyst. The riser temperature increases with the increasing C/O ratio as more heat is brought in by the hot regenerated catalyst.

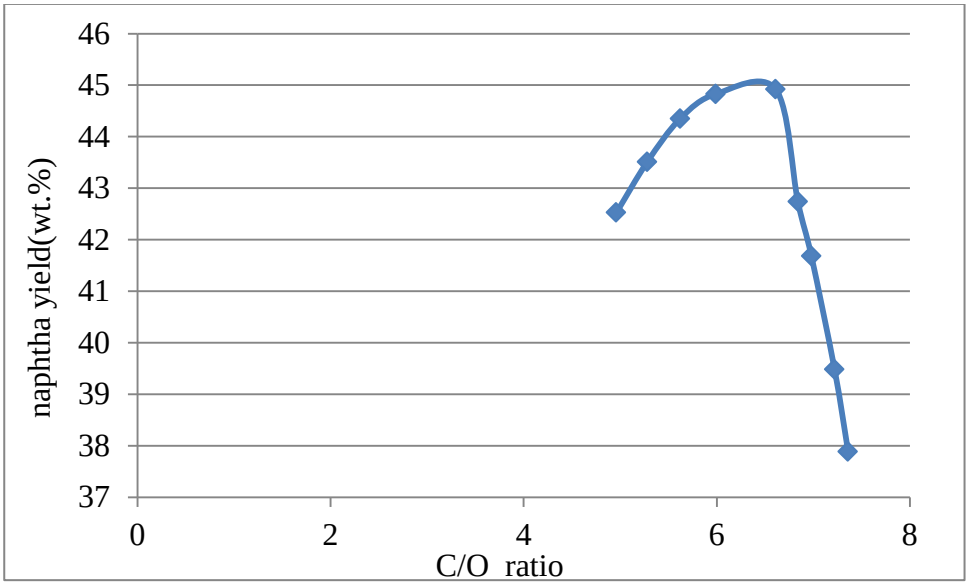

Figure 5: Graph of Naphtha Yield vs. C/O Ratio

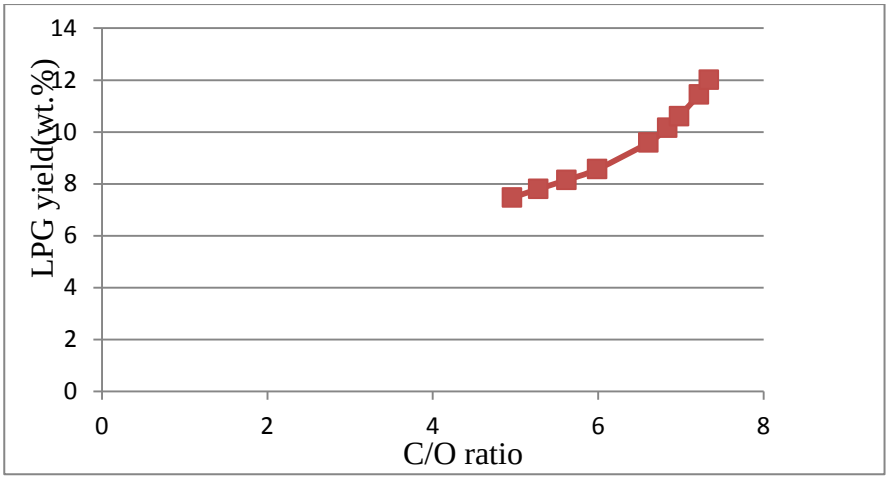

Figure 6: Graph of LPG % vs. C/O Ratio

4.2. Effect of Feed Temperature

Different values of feed temperature were simulated resulting different yield of naphtha and overall conversion. As the temperature of the feed rises from a certain value naphtha yield decreases slightly and so is the total conversion. This is because there is not enough cracking reaction in the riser reactor in presence of the catalyst. Cracking would start before the riser which would decrease the percentage yield of the product.

Table 10: Variation of naphtha & coke yield, total conversion with feed temperature

FEED TEMPERATURE (^{0}C)	NAPHTHA (WT %)	TOTAL CONVERSION (%)	LPG YIELD (WT%)
222.8	39.4548	83.0655	11.4434
232	39.3856	82.8420	11.4559
242	39.2696	82.7038	11.4818
252	39.1455	82.5615	11.5044
262	39.0136	82.4151	11.5256

4.3 Effect of Flowrate

As flow rate of the feed oil to the riser increases, first the naphtha yield increases to a certain point and further increasing the flow rate yield decreases as shown by the graph below. This is because ,with high flow rate riser time decreases resulting less yield of naphtha; and then decreasing flow rate riser time increases which results to more yield. After a certain flow rate the riser time becomes very high resulting more cracking of naphtha to lighter components .but the total conversion increases with increase of the riser time.

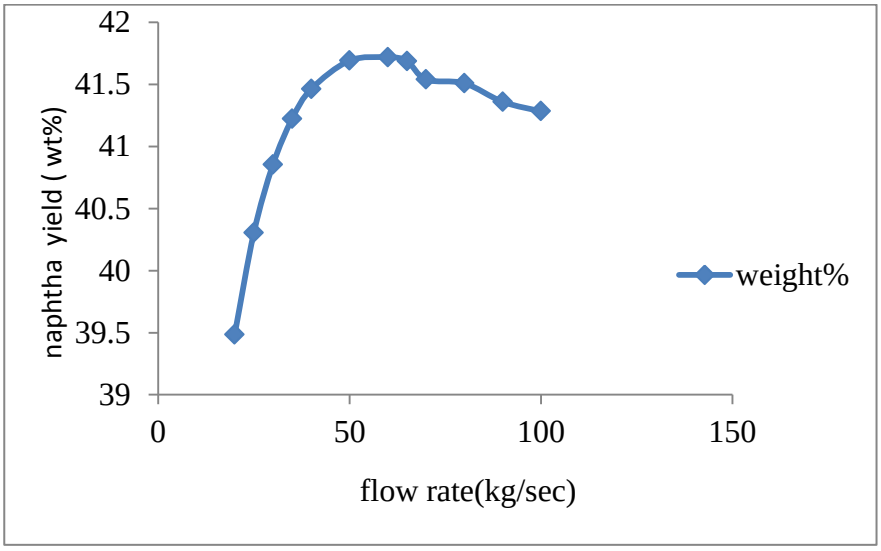

Figure 7: Effect on Naphtha Yield % vs. Feed Flow Rate

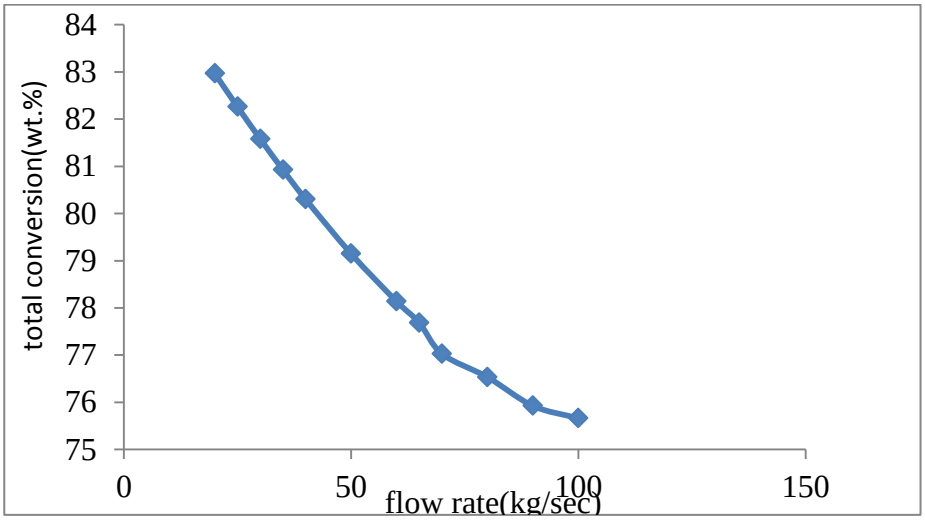

Figure 8: Effect on total Conversion % vs. Feed Flow Rate

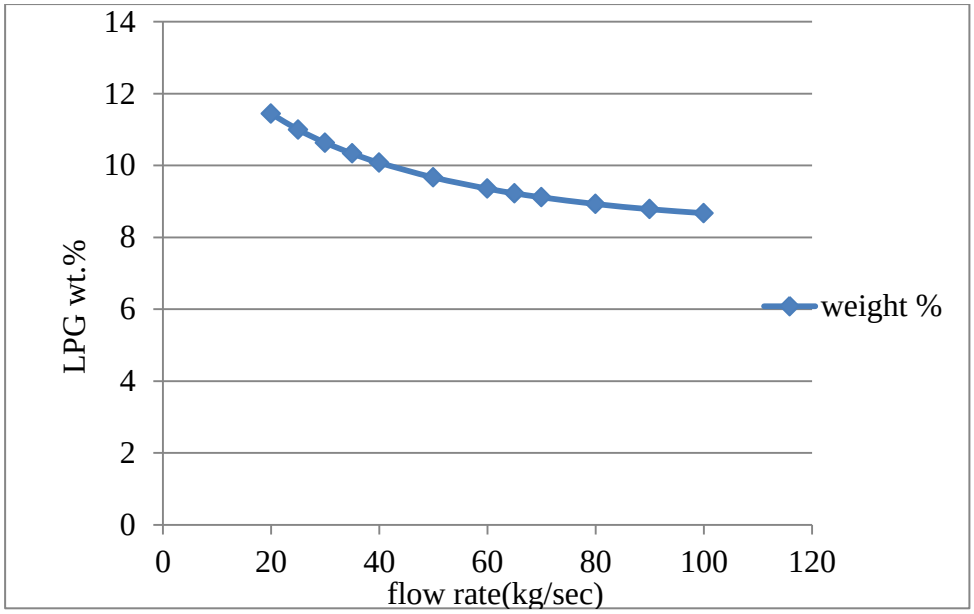

Figure 9: effect of flow rate on LPG yield

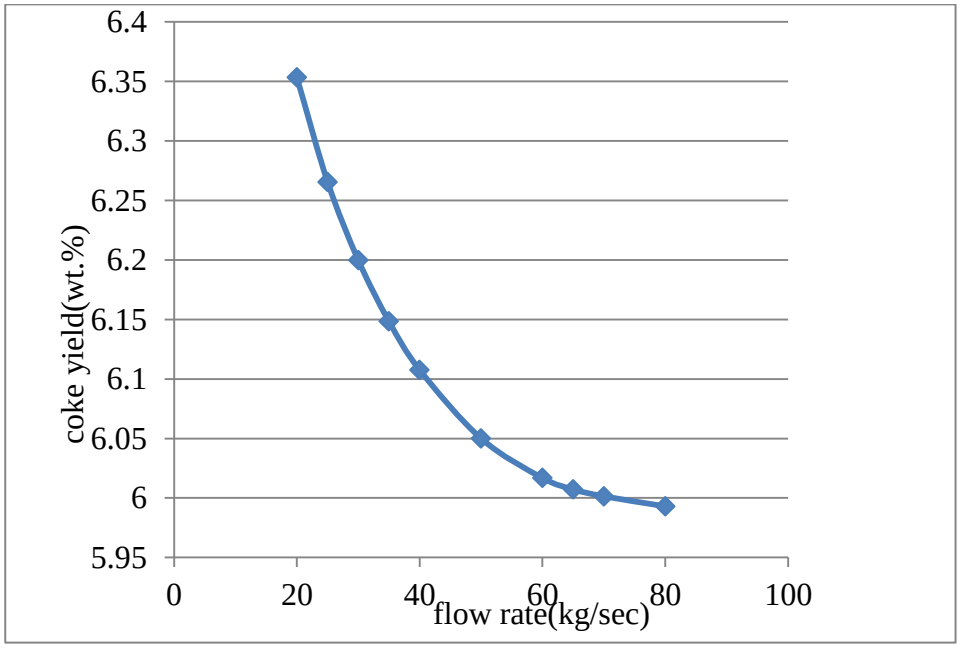

Figure 10: Effect of flow rate on coke yield

4.4. Comparison of One Riser and Dual Riser

Simulation was done using conquest type catalyst (zeolite 24.38 %) in 2 types of riser reactor i.e. one riser reactor and dual riser reactor at process condition as follows: [17]

Height: 33 meter.
Diameter: 8 meter.
Flow rate: 20kg/sec.
Catalyst to oil ratio 7.2
Feed temperature 495.8K
Catalyst used in the process: Conquest 95

Table 11: simulation data of dual risers at given conditions. (265m^3/hr)

Component	Dual riser
H_2S	0.5711
FUEL GAS	2.8448
PROPANE	1.2267
PROPYLENE	4.5457
N-BUTANE	1.0284
I-BUTANE	2.1904
BUTENES	6.3099
NAPHTHA	43.9055
LCO	15.5900
BOTTOMS	15.6143
COKE YIELD	6.1737
TOTAL	100
CONVERSION	68.7957

4.5 Effect of Flow Rate In Both Reactors:

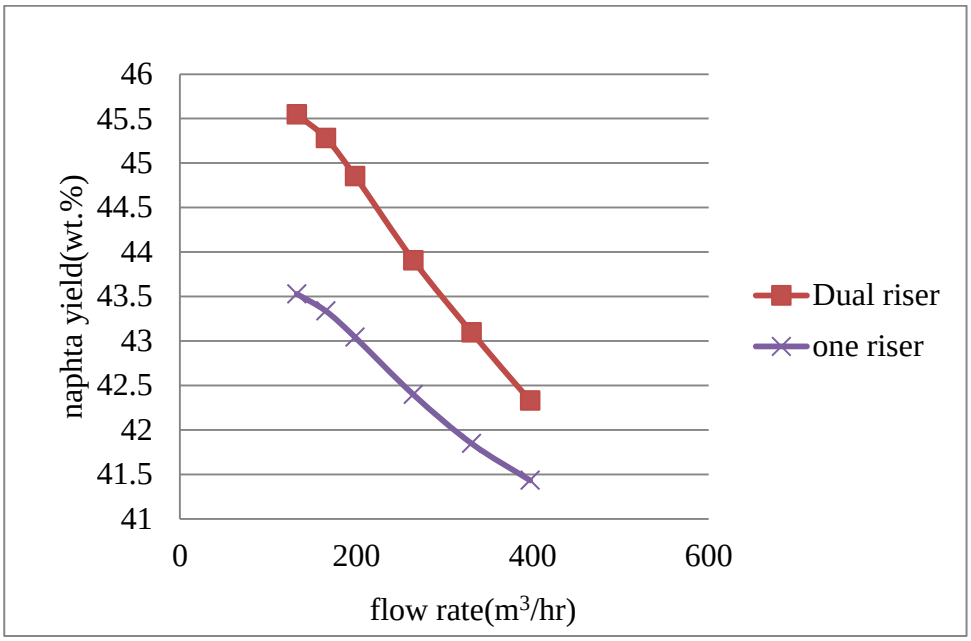

Figure 11: Effect of naphtha yield vs. flow rate

Dual risers reactor are used when there is to maintain maximum flow rate in that case in order to increase the residence time instead of changing the height of the reactor dual riser reactors are used where the steam is divided into two and the flow rate in each riser is the half of the original flow rate. As shown in the graph in between 200m^3/hr. and 600m^3/hr. if flow rate increases the yield decreases. It is same in case of both the reactor. The cause is simple. Due to high flow rate the reaction time in the reactor will be very less, so very less time will be there for efficient contact between catalyst and feed and the naphtha yield decreases as the flow rate increases. At the same flow rate the dual riser shows higher yield than one riser reactor because in case of dual riser the flow rate is divided into two streams, so flow rate will be half and the feed velocity in the riser will be less. So there is efficient time for the cracking process.

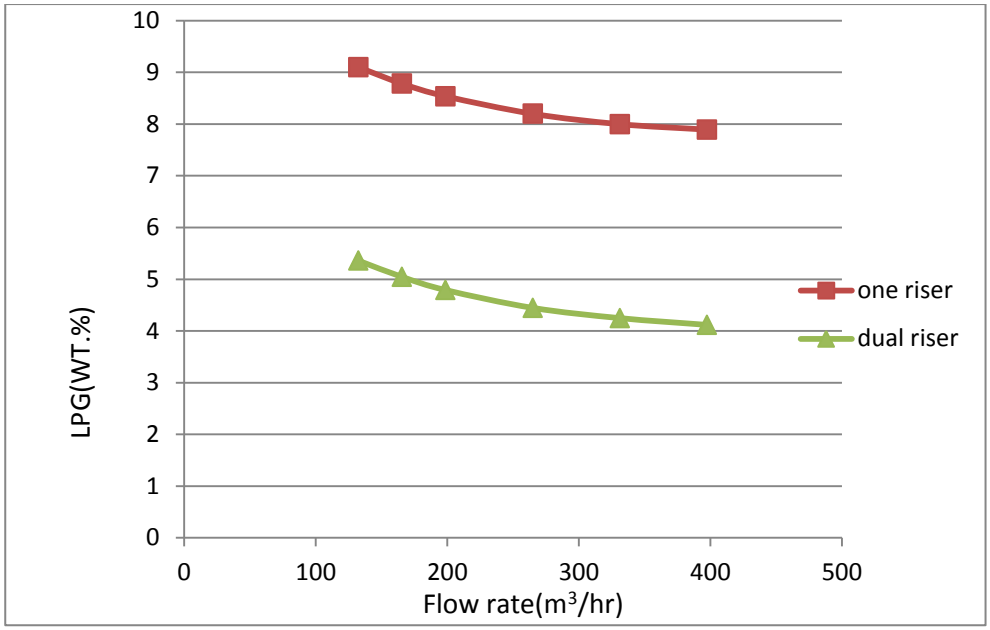

Figure 12 :Effect of flow rate on LPG yield in dual and one riser

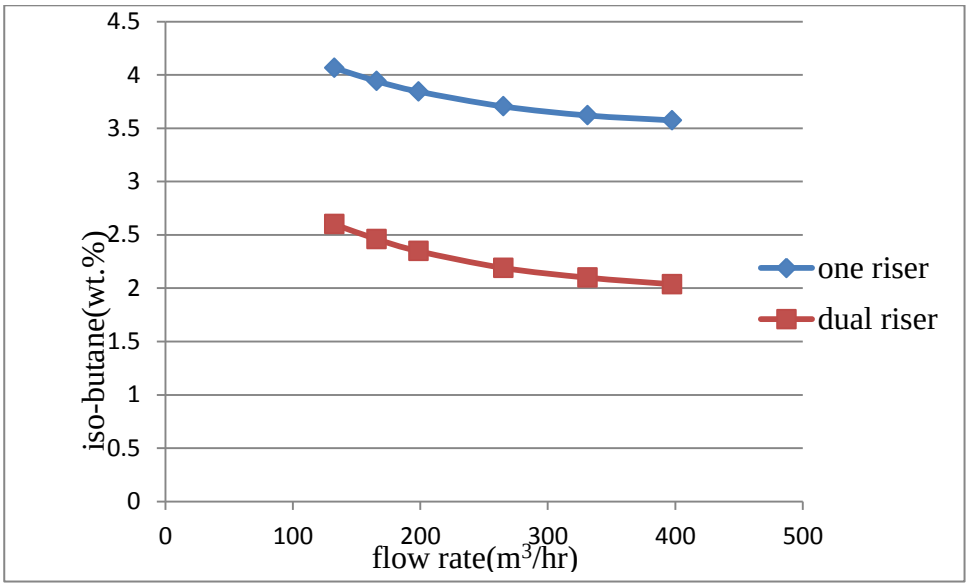

Figure 13: Octane values of naphtha on both the reactors

From the above graph it is obvious that the octane value of naphtha decreases in the dual riser .the graph shows a higher iso-butane content in one riser as compared to dual riser.so yield of naphtha may increase in the dual riser but the octane value of gasoline decreases in major scale.

4.6. Effect of Riser Height

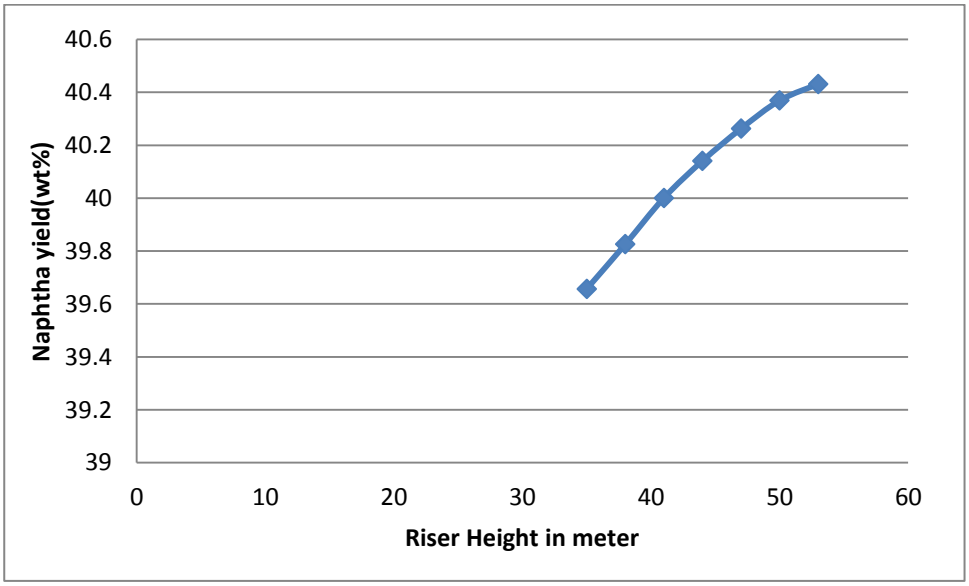

Figure 14: effect of riser height on naphtha yield

From the figure 14 it is known that height is an important parameter in naphtha yield process. By maintaining the same flow rate and reactor pressure and temperature height of the reactor is varied .as shown, naphtha yield will increase as height increases. First it will increase rapidly, but as the height goes on increasing the increase in naphtha yield decreases. The decline in naphtha yield is due to several reasons. As height increases at first the residence time in the reactor increases .this leads to more cracking of the feed .but when height is further increased secondary cracking dominates the process and naphtha yield decreases. In the figure 14: the naphtha yield is still increasing as height increases because the flow rate is maintained at 331m^3/hr. At this flow rate there is minimum residence time in the reactor, so naphtha yield is increasing as height reaches about 60 meters. It can be shown in the table 11 that in case of dual riser at 33meter height and with the same process condition the yield is about 43% which is 39% in case of single riser

4.7 Simulation of the Regenerator:

Simulation of the regenerator is done using ANSYS FLUENT simulation.

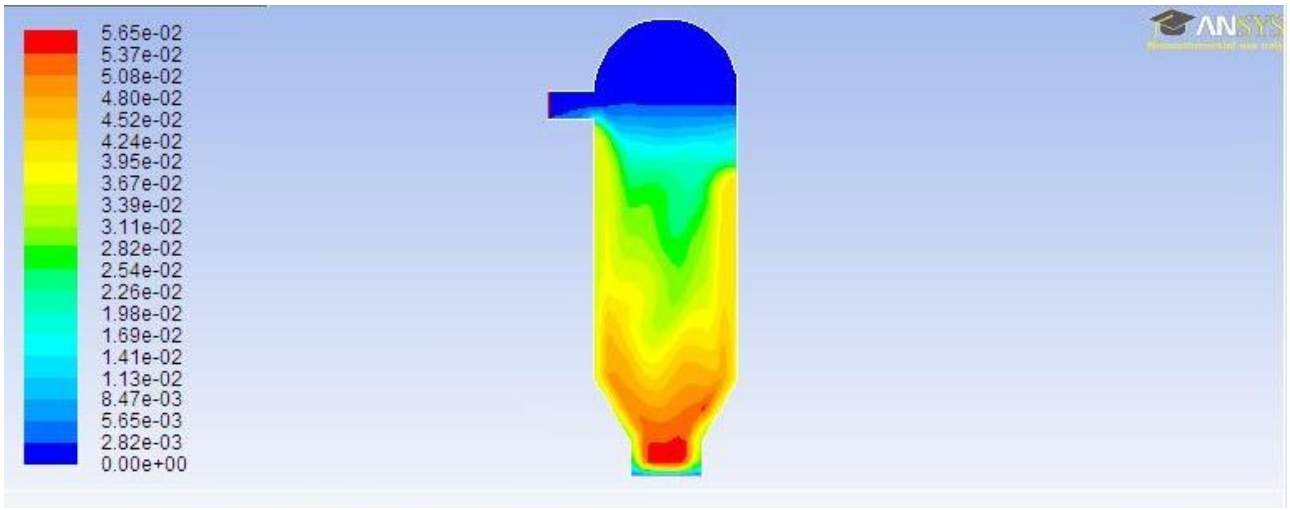

Figure 15: volume fraction distribution of Carbon in regenerator

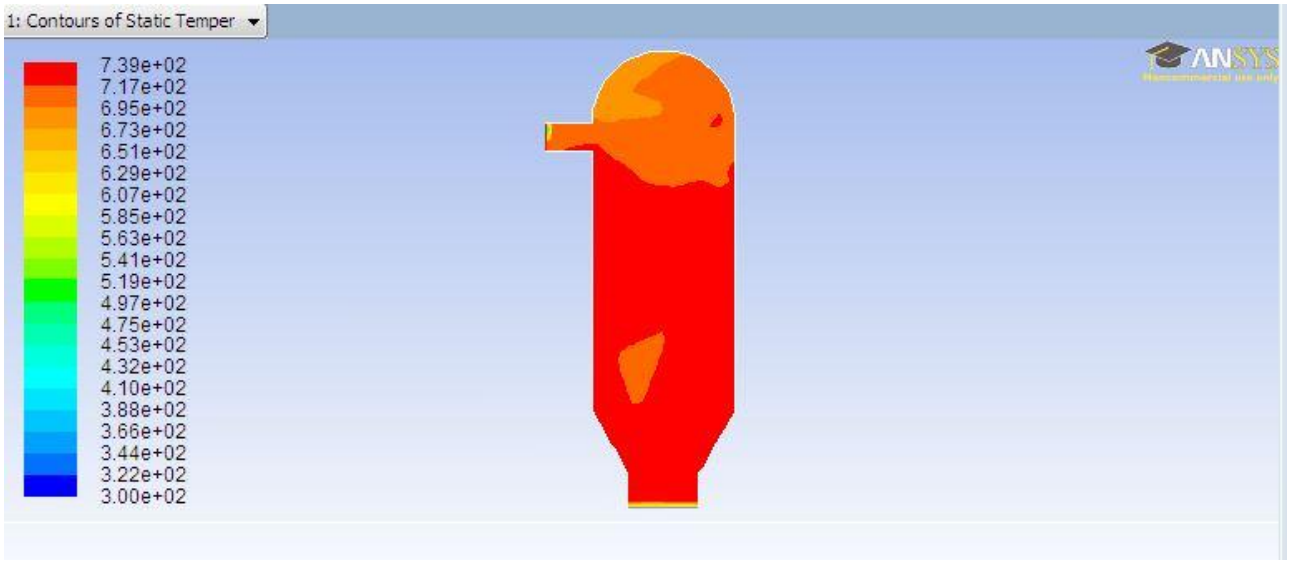

Figure 16: Static temperature Profile in regenerator

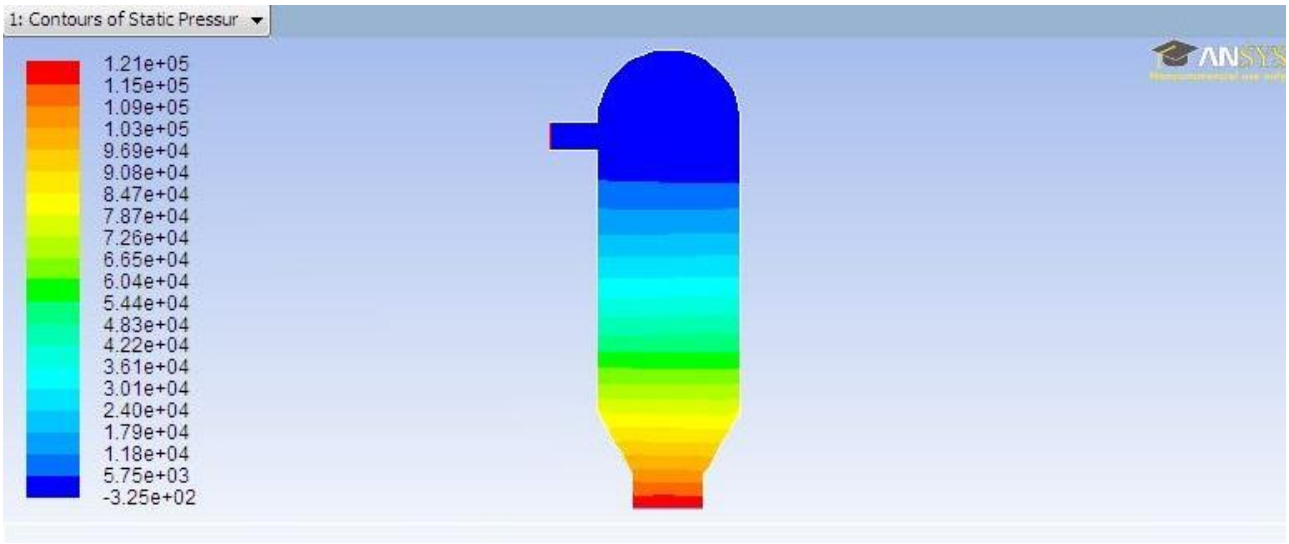

Figure 17: Static Pressure Contour in regenerator

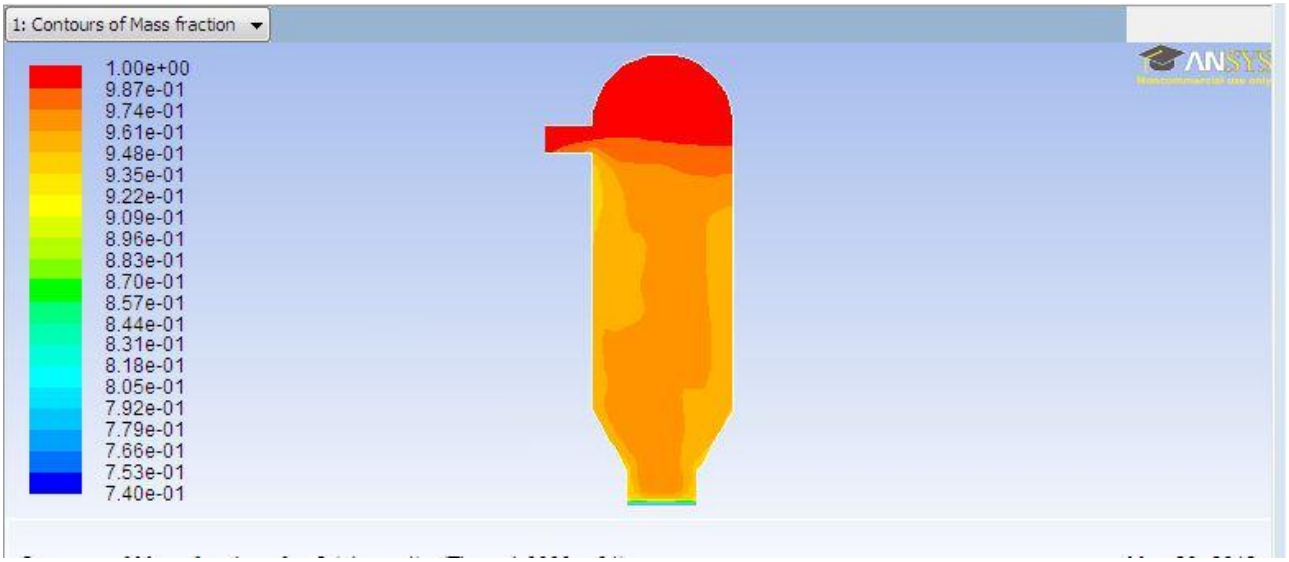

Figure 18: Mass Fraction Distribution of CO_2 in regenerator

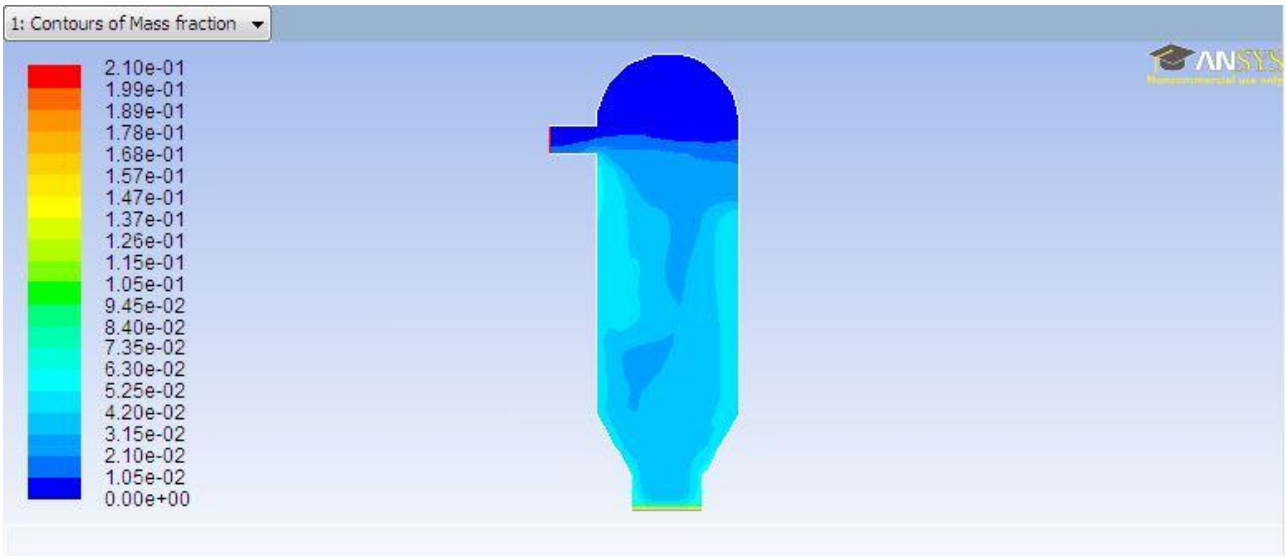

Figure 19: Mass Fraction Distribution of O_2 in regenerator

Simulation of the regenerator is done using velocity inlet as 6 kg /second and pressure outlet as atmospheric pressure and various profiles of the regenerator is studied. In the above figures the temperature, pressure profile are shown. As combustion reaction is occurring in the regenerator the temperature is increased during the process which is shown by the simulation picture. Also the pressure variation is shown. While combustion CO2 is produced which mass fraction distribution is shown in the regenerator simulation picture. Also the velocity inlet O_2 profile is studied and its mass fraction is distributed thorough-out the bed for the combustion purposes. As shown in the figure the oxygen mass fraction in the upper portion of the regenerator is found negligible which shows there is almost complete combustion in the regenerator and in the flue gas oxygen percentage is minimum. So there is maximum mass distribution of carbon dioxide in the upper portion of the regenerator.at first the bed and the carbon temperature was taken 522^0C. Which is increased in due course due to combustion .As shown in the figure after 40 seconds of combustion reaction the maximum temperature in the reactor is found to be 737.5^0 C

Moreover the transient simulation is done for forty seconds and the parameters are taken at that point. The time step which was run during simulation is 0.001. If thus precision is further improved, the simulation result will be more perfect. The temperature profile which is shown in the above figure16. Is governed by Arrhenius equation for combustion reaction. The static pressure shown in the figure 17 explains that the reaction zone is the high pressure zone. The O_2 distribution in the

figure 19 shows complete combustion. This is supported by CO_2 distribution which is one in the upper region of the reactor. All this explains how regenerator is the temperature source for the riser reactor.

Temperature profile at various O_2 concentrations:

Oxygen enrichment is done in various industrial processes in order to perform sufficient combustion. In the batch regenerator combustion is performed at different oxygen concentration. The first case was with no enrichment i.e. 0.21 O_2 concentrations. Then the simulation is run at various level of concentration, 0.15, 0.25, 0.4 .and the temperature profile is noted at 40 seconds. As the carbon content is less and the process is a batch one, so the variation of temperature will be less.

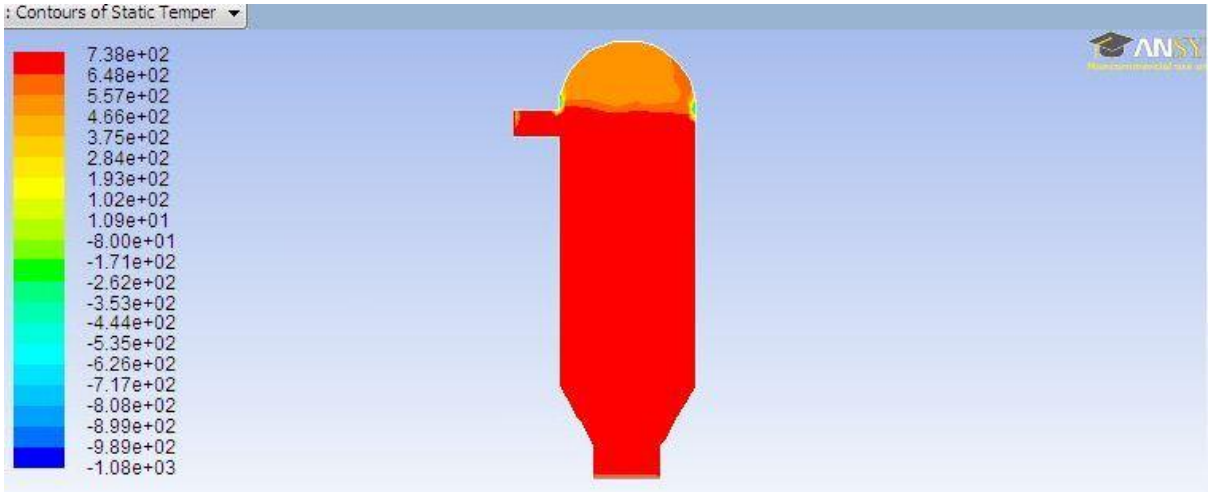

Fig 20: Temperature profile at 0.15 oxygen concentration on air

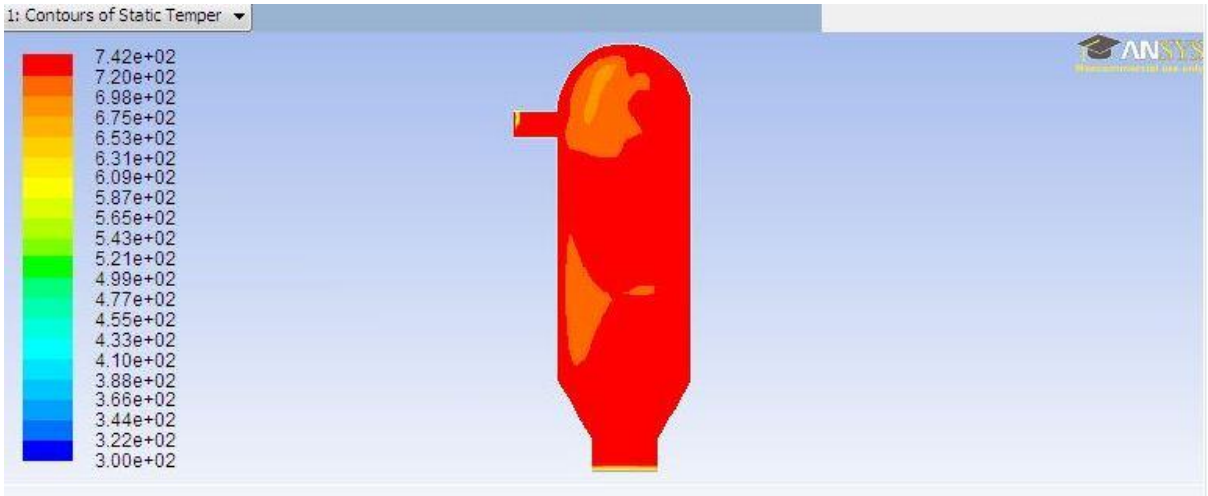

Fig 21: Temperature profile at 0.25 oxygen concentration on air

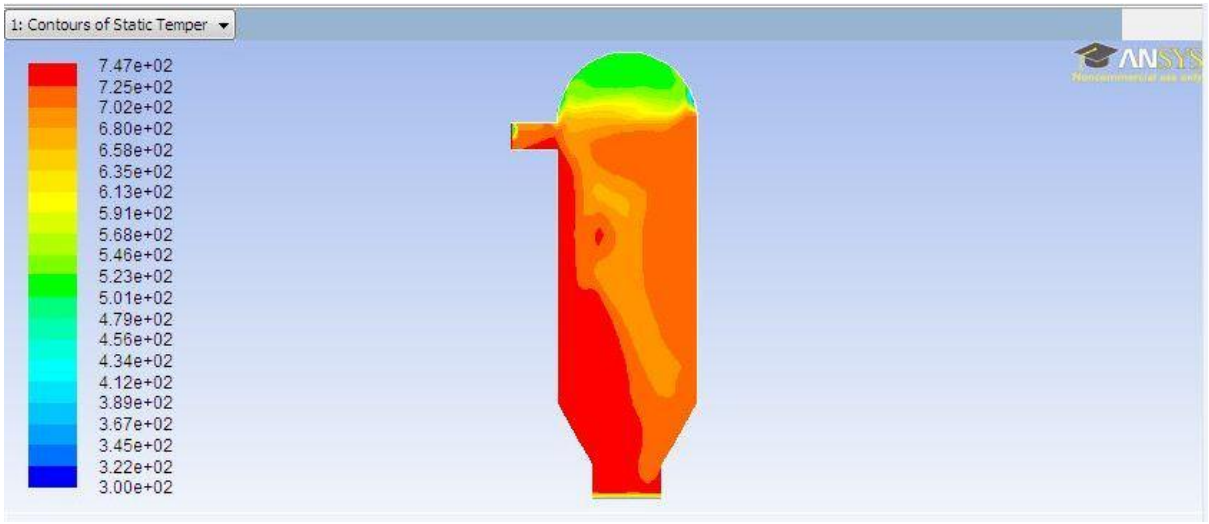

Fig 22: Temperature profile at 0.4 oxygen concentration on air

So in total 4 cases were studied for temperature profile. With increase in oxygen concentration in the feed air the temperature of the regenerator increases. But as oxygen concentration increases to a maximum (in this case 0.4) there is anomaly in the behaviour.it seems that vigorous combustion occurs in some region leading to high temperature profile in some region. So there is channeling of major oxygen through the regenerator to the flue gas. So there exists high temperature gradient.

Temperature profile at different flow rate:

At different flow rate of air, the regenerator is simulated to observe the temperature profile of it. Simulation of these profiles are given below

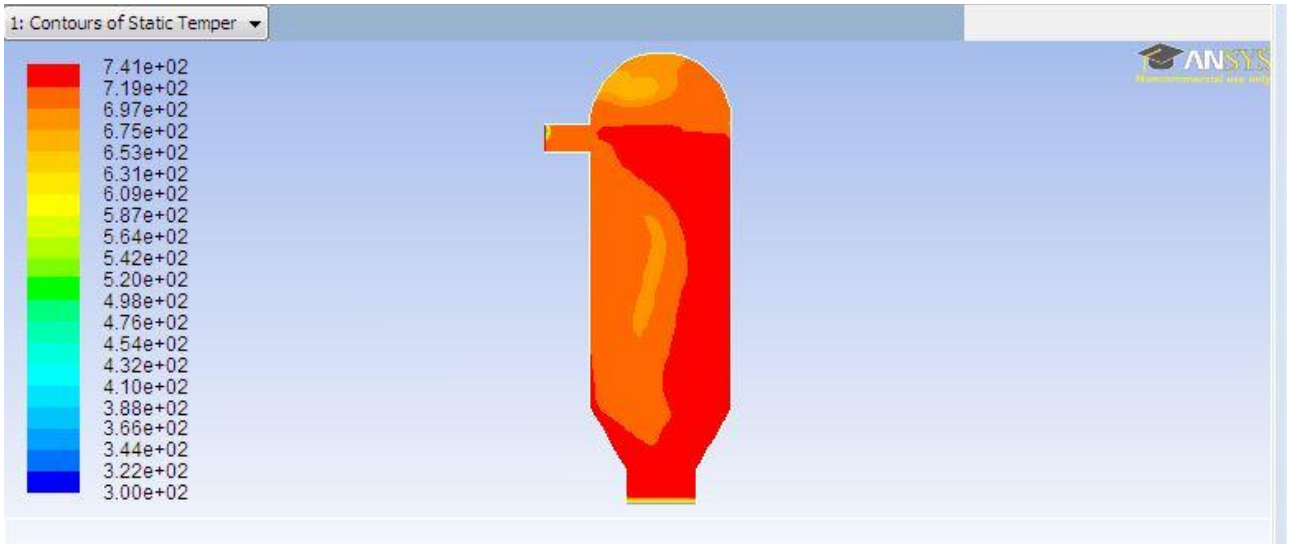

Fig23: temperature profile at flow rate of 7 kg/sec

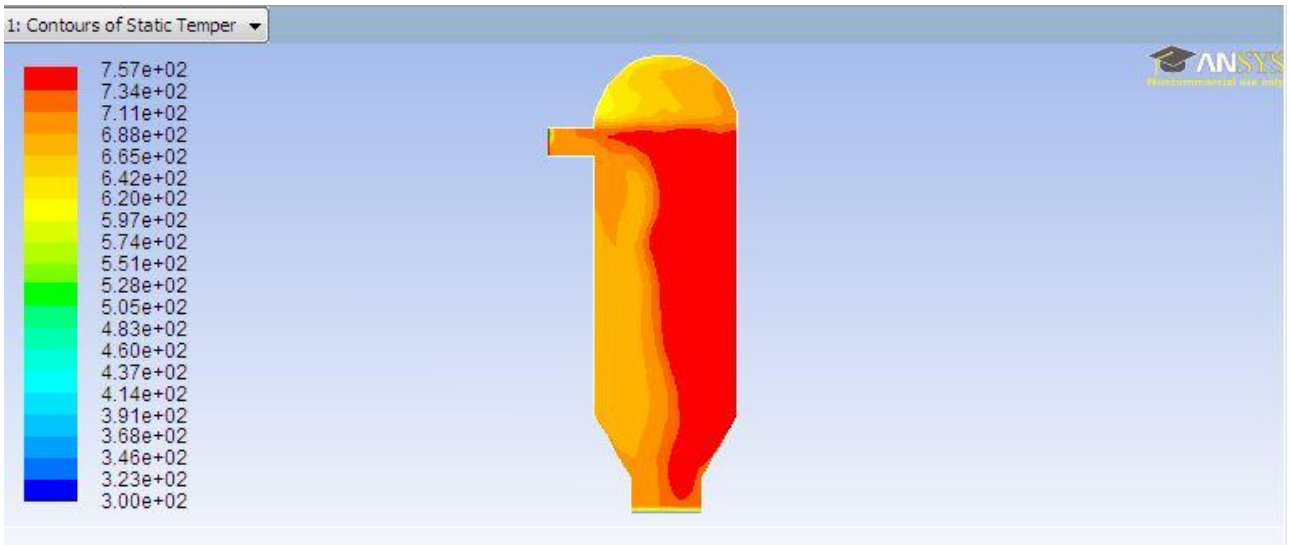

Fig24: temperature profile at flow rate of 10kg/sec

Flow rate was varied and increased to note down the temperature profile inside the regenerator.at the initial case flow rate was maintained at 6 kg/sec and now two additional flow at 7 and at 10 kg/sec was maintained to obtain different simulation result. Finally it was observed that as the flow rate increases the regenerator temperature increases . The rate of increase of temperature increases as flow rate increases. As at 10 kg/sec flow rate the temperature profile of the regenerator is found to be 757^0C. This indicates severe combustion at some zones of the regenerator. But as the flow was increased to a very high level, to an optimum level then the rate of temperature increase decreases gradually. All the flow rate of the air was maintained at the basic 0.21 concentration of O_2 in the supplied air.

4.8 Fractionator simulation:

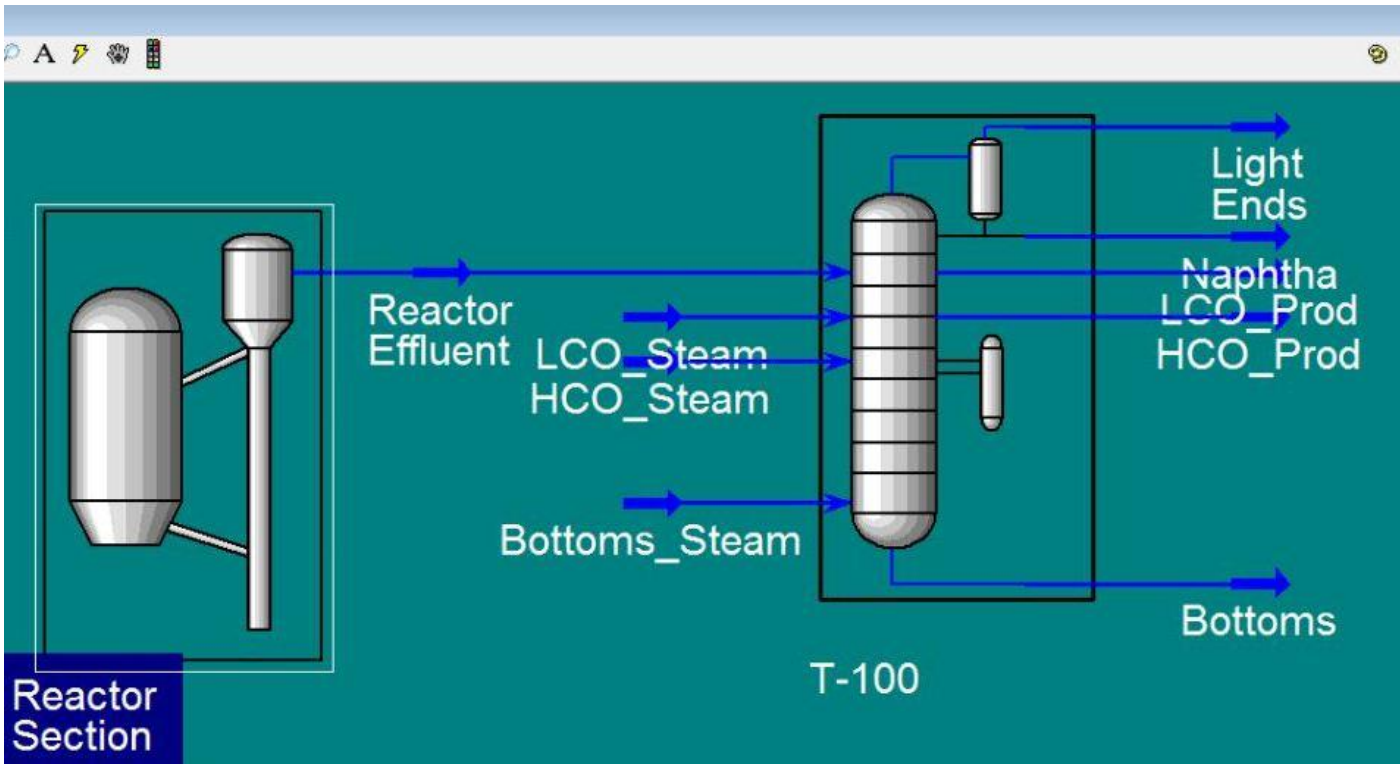

fig25: PFD simulation of FCC fractionator with aspen hysys

Fractionator unit is the separating unit of the FCC. So fractionator simulation is done in order to collect the riser yield separately. Aspen Hysys does the simulation of the FCC unit along the fractionator .The details specification of the procedure is mentioned in the appendix. But some of the process ought to be discussed now. The reactor effluent which comes from the riser separator goes to the fractionating column. The fractionating column has 13 main trays .the reactor effluent which comes to the fractionator is free from the carbon as most of the carbon is deposited on the catalyst surface whish got separated in the regenerator.

The rector effluent comes to the 12^{th} tray of the fractionator and the bottom steam is sent to the bottom most trays. Different zones like bottom zone, HCO zone, LCO zone are there. The LCO zone starts from the condenser to the 4^{th} tray where HCO zone starts from tray 5 to tray 8. The bottom zone starts from 9^{th} to 13^{th} trays. Both LCO and HCO have side strippers which are of 2 stages. The outlet products are light ends, naphtha , where light ends are the over-head vapour products and naphtha is the overhead liquid product. Pressure and temperature maintained

according to atmospheric distillation column. The riser parameter is taken according to Ali et al plat case study. The flow-rate is maintained different.

Feed Flow-rate to the riser: 60m^3/hr.

Catalyst used: AF3 catalyst

The naphtha (gasoline) yield in the above condition is about 42%

Carbon yield is about 5.8%

So in the fractionator, the effluent does not have that carbon .so the mass fraction of naphtha increases.

Table 12: components mass flow ion the fractionator

Components	Mass Flow (kg/hr.)
Reactor Effluent	183855.6
Bottoms Steam	99.868
LCO Steam	99.868
HCO Steam	734.47
Light Ends	46558.43
Naphtha	85071.36
Water	8125.008
Bottoms	6884.1
LCO Prod	19968.7
HCO Prod	12869.3

So fractionator is only the separator unit which does not affect the cracking riser activity. Here no coke is found that's why the ratio to naphtha and reactor effluent is more than the mass fraction of naphtha found on the FCC reactor

5. Conclusion

Simulation of the FCC unit was done and the results of the output are obtained .Naphtha which is a major component in building gasoline, its yield has been obtained (39.485). But in the simulation process inlet feed to the FCC is obtained from the atmospheric distillation. After studying all the process parameters it is observed that there is an optimum condition for naphtha yield. Parameters like flow rate ,C/O ratio ,reactor temperature ,pressure, feed temperature are varied to observe the operation of FCC, Moreover simulation has been at various process conditions using various process parameters, like flow rate, feed and catalyst temperature catalyst to oil ratio and the maximum yield at these processes has been noted down. The most interesting phenomenon is the variation in the composition of the catalyst. In the case of af3 the yield may be less but oil with high octane number is obtained .so it is concluded that the selectivity of the catalyst depends entirely upon the process plant i.e. which is our desired product and accordingly catalyst are used. From the various graphs it is seen that there is an optimum condition for each process and plant should run by it to get the maximum result. The naphtha yield % in dual riser is found to be 43.9055%

Also the ANSYS simulation of the regenerator (for a batch process) is also done and the various parameters profile is obtained. Temperature increase in the regenerator decides how much heat will flow from the regenerator to the riser. Combustion reaction is well established in the simulation and compositions (oxygen, carbon -dioxide, carbon) are shown accordingly. From the oxygen and carbon dioxide mass fraction distribution it is explained that there is complete combustion in the regenerator. The temperature rise is due to that combustion reaction which is governed by Arrhenius equation. The pressure profile is simple and self –explanatory whish shows high pressure at the reaction zone. The regenerator simulation is done for the better understanding of the fluidized catalytic cracking unit and also the riser mechanism.

6. References:

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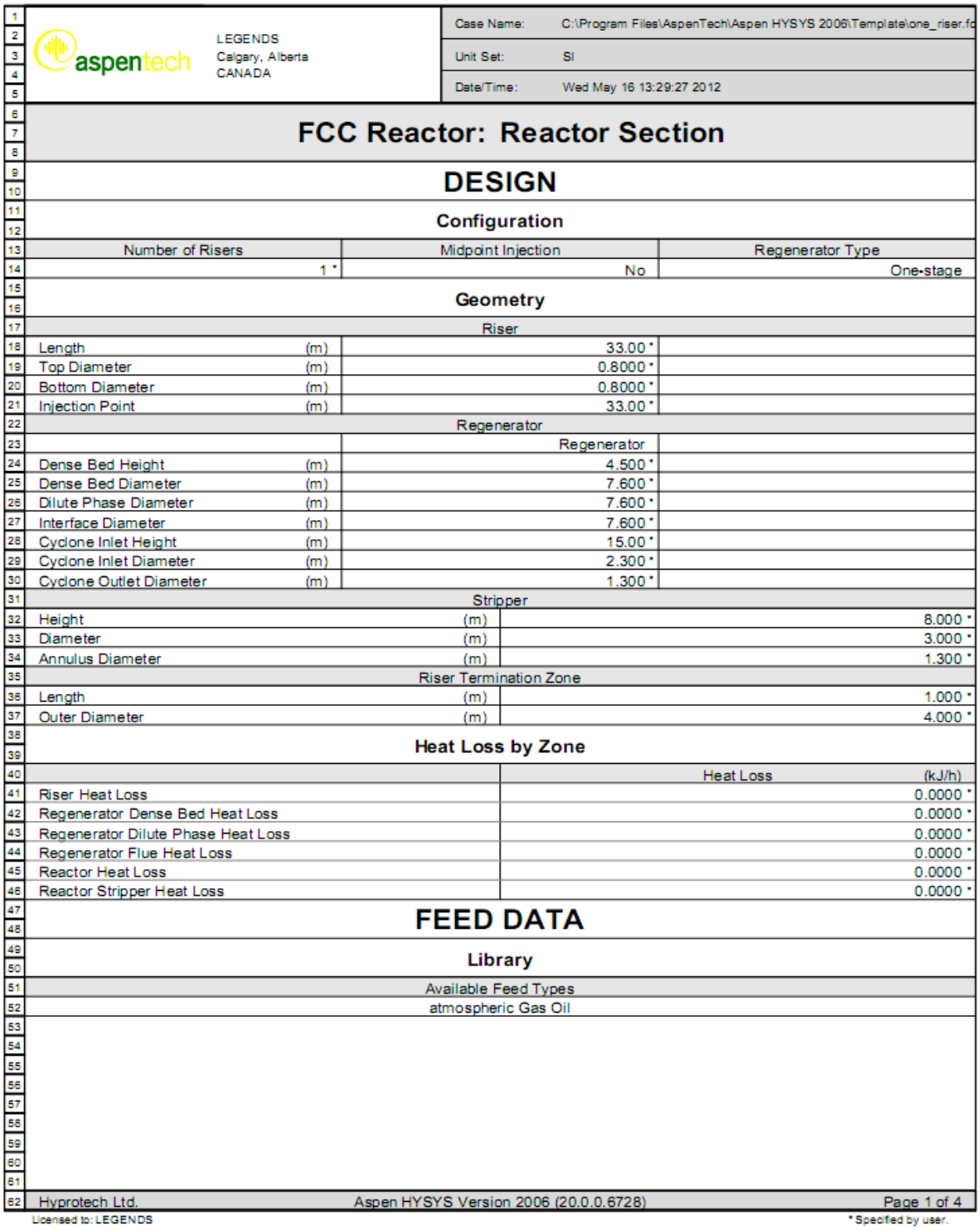
7. Appendix

a) One Riser



LEGENDS
Calgary, Alberta
CANADA

Case Name: C:\Program Files\AspenTech\Aspen HYSYS 2006\Template\one_riser.fc
Unit Set: SI
Date/Time: Wed May 16 13:29:27 2012

FCC Reactor: Reactor Section

DESIGN

Configuration

Number of Risers	Midpoint Injection	Regenerator Type
1 *	No	One-stage

Geometry

Riser

Length	(m)	33.00 *	
Top Diameter	(m)	0.8000 *	
Bottom Diameter	(m)	0.8000 *	
Injection Point	(m)	33.00 *	

Regenerator

		Regenerator	
Dense Bed Height	(m)	4.500 *	
Dense Bed Diameter	(m)	7.600 *	
Dilute Phase Diameter	(m)	7.600 *	
Interface Diameter	(m)	7.600 *	
Cyclone Inlet Height	(m)	15.00 *	
Cyclone Inlet Diameter	(m)	2.300 *	
Cyclone Outlet Diameter	(m)	1.300 *	

Stripper

Height	(m)	8.000 *
Diameter	(m)	3.000 *
Annulus Diameter	(m)	1.300 *

Riser Termination Zone

Length	(m)	1.000 *
Outer Diameter	(m)	4.000 *

Heat Loss by Zone

	Heat Loss	(kJ/h)
Riser Heat Loss		0.0000 *
Regenerator Dense Bed Heat Loss		0.0000 *
Regenerator Dilute Phase Heat Loss		0.0000 *
Regenerator Flue Heat Loss		0.0000 *
Reactor Heat Loss		0.0000 *
Reactor Stripper Heat Loss		0.0000 *

FEED DATA

Library

Available Feed Types
atmospheric Gas Oil

Hyprotech Ltd. Aspen HYSYS Version 2006 (20.0.0.6728) Page 1 of 4
Licensed to: LEGENDS
* Specified by user.

aspentech	LEGENDS Calgary, Alberta CANADA	Case Name:	C:\Program Files\AspenTech\Aspen HYSYS 2006\Template\one_riser.fc
		Unit Set:	SI
		Date/Time:	Wed May 16 13:29:27 2012

FCC Reactor: Reactor Section (continued)

CATALYST

Library

Available Catalysts

A/F-3

Blend

Base Catalyst Blend and Composition

	Weight Fraction	Zeolite	Alumina	Rare Earth
A/F-3	1.0000 *	26.69	37.20	3.746e-002
Total	1.0000	26.69	37.20	3.746e-002

ZSM 5 Additive

Selectivity	Standard
ZSM-5 per unit mass of base blend	0.0000 *

Heat Capacities

Catalyst Heat Capacity	(kJ/kg-C)	1.100 *
Coke Heat Capacity	(kJ/kg-C)	1.670 *

Activity

Feed Metals

	Vanadium (ppmwt)	Nickel (ppmwt)	Sodium (ppmwt)	Iron (ppmwt)	Copper (ppmwt)
Feed-1	0.0000	0.2603	0.0000	0.0000	0.0000
Total	---	---	---	---	---
Bias	-3.372	-1.987	-5.395	-7.193	-8.991e-002
Equilibrium Catalyst	750.0 *	500.0 *	2800 *	4000 *	20.00 *

Other

Fresh Make Up Rate	(kg/h)	323.7
Equilibrium MAT	(%)	68.00 *

OPERATION

Feeds

Feed Conditions

Feed	Volume Flow(m3/h)	Mass Flow (kg/h)	Temperature (C)	Pressure (kPa)	Location
Feed-1	78.09	7.200e+004 *	222.8 *	300.0 *	Riser

Total Feed

	Riser		
Fresh Feed Volume (m3/h)	78.09		
Fresh Feed Mass (kg/h)	7.200e+004		
Total Feed Volume (m3/h)	78.09		
Total Feed Mass (kg/h)	7.200e+004		
Total Feed Preheat Duty (kJ/h)	0.0000 *		
Total Feed Temperature (C)	222.8		

Total Feed Summary

		Total
Fresh Feed Volume	(m3/h)	78.09
Fresh Feed Mass	(kg/h)	7.200e+004
Total Feed Volume	(m3/h)	78.09
Total Feed Mass	(kg/h)	7.200e+004
Total Feed Preheat Duty	(kJ/h)	0.0000

aspentech	LEGENDS Calgary, Alberta CANADA	Case Name:	C:\Program Files\AspenTech\Aspen HYSYS 2006\Template\one_riser.fc
		Unit Set:	SI
		Date/Time:	Wed May 16 13:29:27 2012

FCC Reactor: Reactor Section (continued)

Feeds (continued)

Dispersion Steam			
	Riser		
Steam Mass (kg/h)	1414		
Steam to Total Feed Ratio	2.000e-002 *		
Steam Temperature (C)	200.0 *		
Steam Pressure (kPa)	1000 *		

Riser/Reactor

Riser Temperature Control		
Riser Outlet Temperature	(C)	548.0
Reactor Plenum Temperature	(C)	547.5 *
Catalyst Circulation Rate	(kg/h)	5.110e+005
Cat/Oil Ratio		7.226
Reactor Stripping Zone		
Stripping Steam Mass Rate	(kg/h)	1545
Stripping Steam Temperature	(C)	200.0 *
Stripping Steam Pressure	(kPa)	1000 *
Ratio to Catalyst Circulation Rate		3.000 *

Regenerator

	Regenerator		
Dense Bed Temperature (C)	784.0		
Cyclone Temperature (C)	787.7		
Flue Gas Temperature (C)	787.7		
Flue Gas-Dense Bed Delta-T	3.746		
Flue Gas O2 (%)	1.00 *		
Flue Gas CO (%)	0.00		
Flue Gas CO2 (%)	16.76		
Flue Gas CO/CO2 Ratio (%)	0.00		
Carbon on Regen Cat (CRC) (%)	0.01		
Catalyst Cooler Duty (kJ/h)	0.0000 *		
Dense Bed Bulk Density	540.0 *		
Catalyst Inventory (kg)	1.102e+005		
Air Volume Flow (m3/h)	5.162e+004		
Air Mass Flow (kg/h)	6.236e+004		
Enriched O2 Volume Flow (m3/h)	0.0000 *		
Enriched O2 Mass Flow (kg/h)	0.0000		
Enriched O2 Pressure (kPa)	101.0 *		
Enriched O2 Temperature (C)	100.0 *		
Air Blower Discharge Temp (C)	200.0 *		

Ambient Air Box		
Temperature (C)	Pressure (kPa)	Relative Humidity (%)
200.0 *	101.0 *	70.00 *

Stage 1 Conditions	
Dense Bed Temperature	
Apparent	
Bias	
CRC	
Apparent	

aspentech	LEGENDS Calgary, Alberta CANADA	Case Name:	C:\Program Files\AspenTech\Aspen HYSYS 2006\Template\one_riser.fc
		Unit Set:	SI
		Date/Time:	Wed May 16 13:29:27 2012

FCC Reactor: Reactor Section (continued)

Regenerator (continued)

Bias		

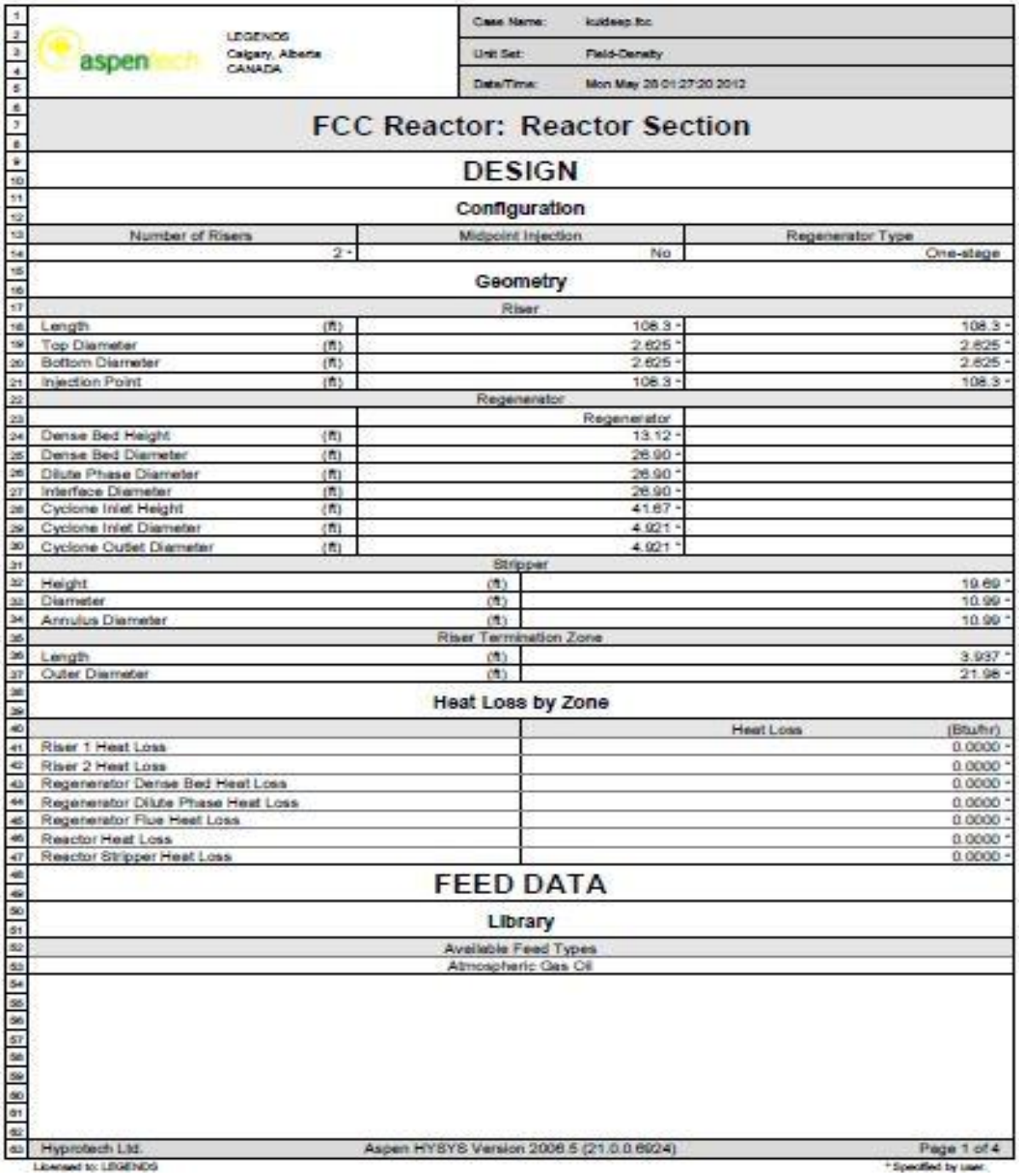
Pressure Control

Reactor Pressure	(kPa)	340.0 *
Regenerator Stage 1 Pressure	(kPa)	370.0
Regenerator Stage 2 Pressure		
Regenerator Stage 2 - Reactor Pressure Difference	(kPa)	30.00 *
Regenerator Stage 2 - Riser Pressure Difference	(kPa)	32.05

Solver Options

Iteration Limits	
Maximum Iterations	Minimum Iterations
20.00 *	0.0000 *
Convergence Tolerance	
Residual	1.000e-006 *
Variable Scaling Parameter	
On/Off Switch	On
Failure Recovery Action	
Action	Revert to the previous results
Creep Step Parameters	
On/Off Switch	Off
Iterations	10.00 *
Step Size	0.1000 *
SQP Hessian Parameters	
Initialization	Normal
Scaling Factor	1.000 *
Updates stored	10.00 *
Linear Search Parameters	
Algorithm	Normal
Step Control	Normal
Step Control Iterations	0.0000 *

b) Dual Riser



LEGENDS
Calgary, Alberta
CANADA

Case Name: kuldeep.ibc
Unit Set: Field-Density
Date/Time: Mon May 28 01:27:20 2012

FCC Reactor: Reactor Section

DESIGN

Configuration

Number of Risers	Midpoint Injection	Regenerator Type
2 *	No	One-stage

Geometry

Riser

Length	(ft)	108.3 *	108.3 *
Top Diameter	(ft)	2.625 *	2.625 *
Bottom Diameter	(ft)	2.625 *	2.625 *
Injection Point	(ft)	108.3 *	108.3 *

Regenerator

		Regenerator
Dense Bed Height	(ft)	13.12 *
Dense Bed Diameter	(ft)	26.90 *
Dilute Phase Diameter	(ft)	26.90 *
Interface Diameter	(ft)	26.90 *
Cyclone Inlet Height	(ft)	41.67 *
Cyclone Inlet Diameter	(ft)	4.921 *
Cyclone Outlet Diameter	(ft)	4.921 *

Stripper

Height	(ft)	19.69 *
Diameter	(ft)	10.99 *
Annulus Diameter	(ft)	10.99 *

Riser Termination Zone

Length	(ft)	3.937 *
Outer Diameter	(ft)	21.98 *

Heat Loss by Zone

	Heat Loss	(Btu/hr)
Riser 1 Heat Loss		0.0000 *
Riser 2 Heat Loss		0.0000 *
Regenerator Dense Bed Heat Loss		0.0000 *
Regenerator Dilute Phase Heat Loss		0.0000 *
Regenerator Flue Heat Loss		0.0000 *
Reactor Heat Loss		0.0000 *
Reactor Stripper Heat Loss		0.0000 *

FEED DATA

Library

Available Feed Types
Atmospheric Gas Oil

Hyprotech Ltd. Aspen HYSYS Version 2006.5 (21.0.0.6924) Page 1 of 4
Licensed to: LEGENDS
* Specified by user.

aspentech	LEGENDS Calgary, Alberta CANADA	Case Name:	koldeep.hc
		Unit Set:	Field-Density
		Date/Time:	Mon May 28 01:27:20 2012

FCC Reactor: Reactor Section (continued)

CATALYST

Library

Available Catalysts
Conquest 95

Blend

Base Catalyst Blend and Composition

	Weight Fraction	Zeolite	Alumina	Rare Earth
Conquest 95	1.0000 *	24.39	39.69	12.01
Total	1.0000	24.39	39.69	12.01

ZSM 5 Additive

Selectivity	Standard
ZSM-5 per unit mass of base blend	0.0000 *

Heat Capacities

Catalyst Heat Capacity	(Btu/lb-F)	0.2627 *
Coke Heat Capacity	(Btu/lb-F)	0.3949 *

Activity

Feed Metals

	Vanadium (ppmwt)	Nickel (ppmwt)	Sodium (ppmwt)	Iron (ppmwt)	Copper (ppmwt)
Feed-1	0.0000	0.2603	0.0000	0.0000	0.0000
Total	0.0000	0.2603	0.0000	0.0000	0.0000
Bias	-5.684	-3.516	-5.286	-11.33	-0.1510
Equilibrium Catalyst	750.0 *	500.0 *	2800 *	4000 *	20.00 *

Other

Fresh Make Up Rate	(lb/hr)	3050
Equilibrium MAT	(%)	68.00 *

OPERATION

Feeds

Feed Conditions

Feed	Volume Flow (barrel/day)	Mass Flow (lb/hr)	Temperature (F)	Pressure (psia)	Location
Feed-1	3.000e+004 *	4.039e+005	433.0 *	43.51 *	Split

Total Feed

	Riser 1	Riser 2	
Fresh Feed Volume (barrel/day)	1.584e+004	1.416e+004	
Fresh Feed Mass (lb/hr)	2.133e+005	1.907e+005	
Total Feed Volume (barrel/day)	1.584e+004	1.416e+004	
Total Feed Mass (lb/hr)	2.133e+005	1.907e+005	
Total Feed Preheat Duty (Btu/hr)	0.0000 *	0.0000 *	
Total Feed Temperature (F)	433.0	433.0	

Total Feed Summary

	Total
Fresh Feed Volume (barrel/day)	3.000e+004
Fresh Feed Mass (lb/hr)	4.039e+005
Total Feed Volume (barrel/day)	3.000e+004
Total Feed Mass (lb/hr)	4.039e+005
Total Feed Preheat Duty (Btu/hr)	0.0000

Dispersion Steam

	Riser 1	Riser 2	

aspentech	LEGENDS Calgary, Alberta CANADA	Case Name:	kuldeep.hsc	
		Unit Set:	Field-Density	
		Date/Time:	Mon May 28 01:27:20 2012	

FCC Reactor: Reactor Section (continued)

Feeds (continued)

Dispersion Steam (continued)				
Steam Mass	(lb/hr)	6381	5686	
Steam to Total Feed Ratio		3.000e-002 *	3.000e-002 *	
Steam Temperature	(F)	392.0 *	392.0 *	
Steam Pressure	(psia)	145.0 *	145.0 *	

Riser/Reactor

Riser Temperature Control			
Riser Outlet Temperature	(F)		900.0 *
Reactor Plenum Temperature	(F)		893.3
Catalyst Circulation Rate	(lb/hr)		3.115e+006
Cat/Oil Ratio			7.676
Reactor Stripping Zone			
Stripping Steam Mass Rate	(lb/hr)		9344
Stripping Steam Temperature	(F)		392.0 *
Stripping Steam Pressure	(psia)		145.0 *
Ratio to Catalyst Circulation Rate			3.000 *

Regenerator

	Regenerator		
Dense Bed Temperature (F)	1225		
Cyclone Temperature (F)	1239		
Flue Gas Temperature (F)	1239		
Flue Gas-Dense Bed Delta-T	7.653		
Flue Gas O2 (%)	1.00 *		
Flue Gas CO (%)	0.09		
Flue Gas CO2 (%)	16.53		
Flue Gas CO/CO2 Ratio (%)	0.01		
Carbon on Regen Cat (CRC) (%)	0.16		
Catalyst Cooler Duty (Btu/hr)	0.0000 *		
Dense Bed Bulk Density	400.0 *		
Catalyst Inventory (lb)	1.883e+005		
Air Volume Flow (barrel/day)	1.540e+007		
Air Mass Flow (lb/hr)	2.729e+005		
Enriched O2 Volume Flow (barrel/day)	0.0000 *		
Enriched O2 Mass Flow (lb/hr)	0.0000		
Enriched O2 Pressure (psia)	14.65 *		
Enriched O2 Temperature (F)	212.0 *		
Air Blower Discharge Temp (F)	284.0 *		

Ambient Air Box					
Temperature	(F)	Pressure	(psia)	Relative Humidity	(%)
284.0 *		14.65 *		90.00 *	

Stage 1 Conditions	
Dense Bed Temperature	
Apparent	
Bias	
CRC	
Apparent	
Bias	

LEGENDS Calgary, Alberta CANADA		Case Name:	koldeep.hc
		Unit Set:	Field-Deraity
		Date/Time:	Mon May 28 01:27:20 2012

FCC Reactor: Reactor Section (continued)

Pressure Control

Reactor Pressure (psia)		23.93 *
Regenerator Stage 1 Pressure (psia)		34.81 *
Regenerator Stage 2 Pressure		
Regenerator Stage 2 - Reactor Pressure Difference (psi)		10.88
Regenerator Stage 2 - Riser Pressure Difference (psi)		9.995

Solver Options

Iteration Limits	
Maximum Iterations	Minimum Iterations
20.00 *	0.0000 *
Convergence Tolerance	
Residual	1.000e-006 *
Variable Scaling Parameter	
On/Off Switch	On
Failure Recovery Action	
Action	Revert to the previous results
Creep Step Parameters	
On/Off Switch	Off
Iterations	10.00 *
Step Size	0.1000 *
SQP Hessian Parameters	
Initialization	Normal
Scaling Factor	1.000 *
Updates stored	10.00 *
Linear Search Parameters	
Algorithm	Normal
Step Control	Normal
Step Control Iterations	0.0000 *

Af3 catalyst used in FCC:

FCC Catalyst Name	A/F-3				2M1Butene	1.058146
Description	Akzo A/F-3				C2Pentene	0.938267
Created	Oct-20	2003	17:24 17:24:55		T2Pentene	0.957186
Modified	Oct-20	2003	17:24 17:24:55		Cyclopentene	1.046789
Manufacturer	Akzo				Isoprene	0.958755
Kinetic Coke	1.045989				Benzene	1.5625
Feed Coke	1.166873				Metals H2	1.563636
Stripping Eff.	0.999811				Heat Of Rxn.	0
Metals Coke	1.057143				Bot. Cracking	-0.03785
Methane	1.307692				Fresh MAT	76.05
Ethylene	1.489796				HT Deact.	1.006145
Ethane	1.121951				Met. Deact.	0.611945
Propylene	1.351955				LN RON	2.412
Propane	1.517483				LN MON	1.194
IC4	1.27598				LN Nap.	-0.34
Total C4=	1.318519				LN Olefins	7.28
N Butane	1.051095				LN Aromatics	1.155
IC5	1.235693				LCO SPGR	-0.00837
Total C5=	1.38799				CSO SPGR	-0.0091
NC5	1.017909				SOx	1.037847
IC4=	1.189059				HN RON	2.377714
1Butene	0.943844				HN MON	1.211143
C2Butene	0.947135				HN Nap.	-0.895
Butadiene	1.398742				HN Olefins	1.337143
Cyclopentane	0.793549				HN Aromatics	7.283571
3M1Butene	1.052484				LN SPGR	0.005483
1Pentene	0.92546				HN SPGR	0.007414

Af3 catalyst:

Spare 50	0
ZSA M2/GM	166.8
MSA M2/GM	174.8
Zeolite(Wt%)	26.694407
Alumina(Wt%)	37.2
ZRE(Wt%)	0.037461
Sodium(ppm)	1600
Nickel(ppm)	0
Vanadium(ppm)	0
Copper(ppm)	0
Iron(ppm)	2400
ZSM5 LN RON	0
ZSM5 LN MON	0
ZSM5 HN RON	0
ZSM5 HN MON	0
Price	0
Spare 66	0
Spare 67	0
Spare 68	0
Spare 69	0
Spare 70	0

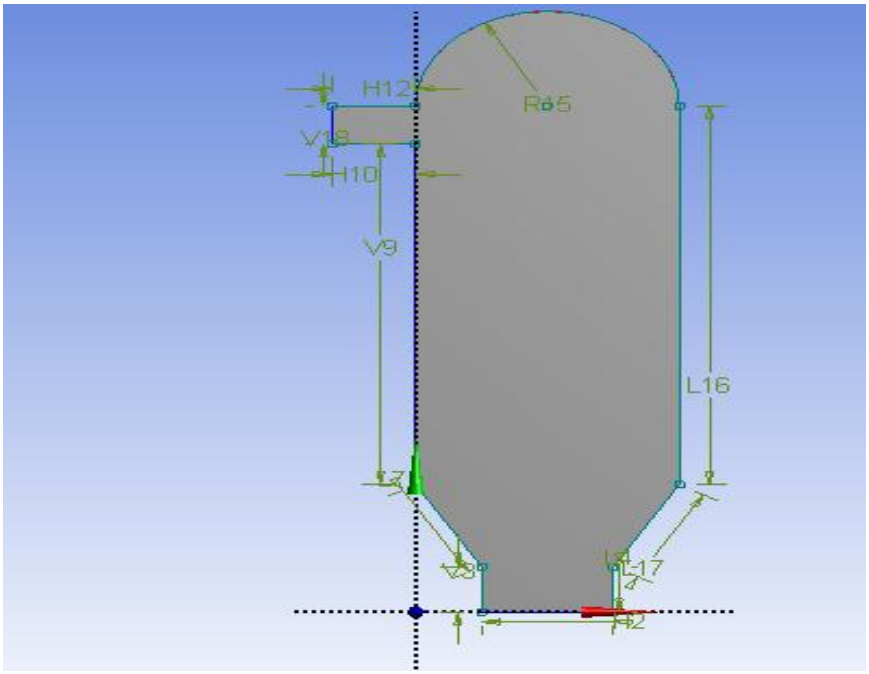
Conquest 95 catalyst used in FCC

FCC Catalyst Name	Conquest 95					
Description	Akzo Conquest 95					
Created	Oct-20	2003	17:40 17:40:42		2M1Butene	1
Modified	Oct-20	2003	17:40 17:40:42		C2Pentene	1
Manufacturer	Akzo				T2Pentene	1
Kinetic Coke	1				Cyclopentene	1
Feed Coke	1				Isoprene	1
Stripping Eff.	1				Benzene	1
Metals Coke	1				Metals H2	1
Methane	1				Heat Of Rxn.	0
Ethylene	1				Bot. Cracking	0
Ethane	1				Fresh MAT	80.8
Propylene	1				HT Deact.	0.5
Propane	1				Met. Deact.	0.5
IC4	1				LN RON	0
Total C4=	1				LN MON	0
N Butane	1				LN Nap.	0
IC5	1				LN Olefins	0
Total C5=	1				LN Aromatics	0
NC5	1				LCO SPGR	0
IC4=	1				CSO SPGR	0
1Butene	1				SOx	1
C2Butene	1				HN RON	0
Butadiene	1				HN MON	0
Cyclopentane	1				HN Nap.	0
3M1Butene	1				HN Olefins	0
1Pentene	1				HN Aromatics	0

Conquest95 catalyst used in FCC

LN SPGR	0
HN SPGR	0
Spare 50	0
ZSA M2/GM	141.7
MSA M2/GM	183.3
Zeolite(Wt%)	24.38689
Alumina(Wt%)	39.69
ZRE(Wt%)	12.01465
Sodium(ppm)	2100
Nickel(ppm)	0
Vanadium(ppm)	0
Copper(ppm)	0
Iron(ppm)	2500
ZSM5 LN RON	0
ZSM5 LN MON	0
ZSM5 HN RON	0
ZSM5 HN MON	0
Price	0
Spare 66	0
Spare 67	0
Spare 68	0
Spare 69	0
Spare 70	0

Regenerator



Dimension in which the regenerator is modelled is shown below using Ai et al case study.

H10 -8ft

H12 -8ft

H2 -12.5ft

L16 -50ft

L17 -12.45ft

R15 -12.5ft

V18 -5ft

V3 -6ft

V9 -35ft

Regenerator after catalyst is patched.